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FLUID BED INCINERATION OF PETROLEUM REFINERY WASTES

American Oil Company
Mandan, North Dakota

March 1971

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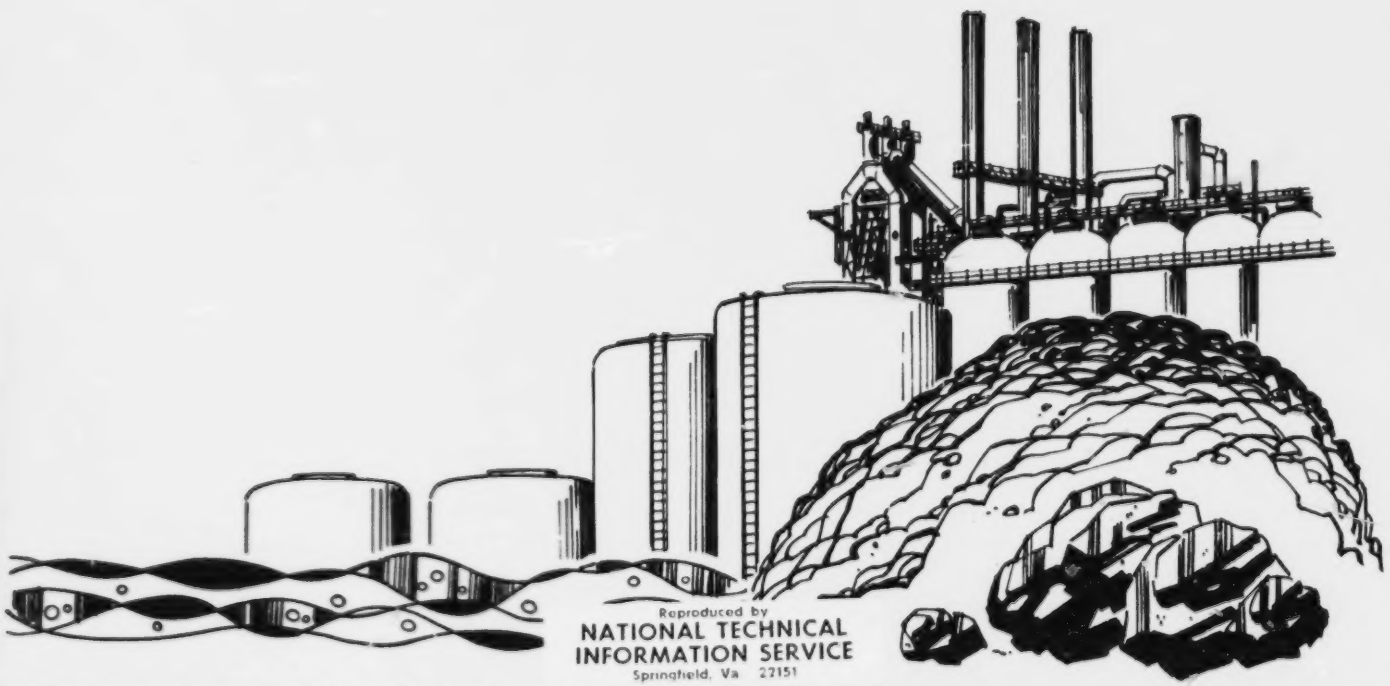

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SELECTED WATER RESOURCES ABSTRACTS
INPUT TRANSACTION FORM

1. Report No. | 2. | 3. Accession No. W

4. Title FLUID BED INCINERATION OF PETROLEUM REFINERY WASTES

5. Report Date

6.

7. Author(s) ~~Simons, Herbert E.~~

8. Performing Organization Report No.

9. Organization American Oil Company, Mandan, North Dakota, Mandan Refinery

10. Project No. 12050 EKT

11. ~~Contract~~/Grant No. 215-01-68

12. Sponsoring Organization

13. Type of Report and Period Covered

15. Supplementary Notes

Water Pollution Control Research Series, ~~No. 12050 EKT, 03/71,~~ 110 ~~pages~~, 11 fig, 9 tab, 2 ref

16. Abstract A fluid bed incinerator was designed for the Mandan Refinery of American Oil Company, by the Copeland Process Corporation, to dispose of 10 Bbl/day oil-water separator sludge, 10 Bbl/day emulsion treating sludge and 10 Bbl/day of spent caustic during a 6 hour day for 9 months of the year. During the 3 remaining months an additional 10 Bbl/day of tank washings was expected, and the incinerator was expected to operate 24 hours/day. The unit which was finally installed was equipped with supplemental fuel lines for start-up, waste feed injection from the top to cool the effluent fuel gases, pressure taps for operational control, and an 80 Bbl sump for mixing. Since the rate at which spent caustic can be incinerated is dependent upon the heat content of the oily sludge, initial attempts to mix the two before injection were abandoned in favor of separate feed lines. When injected above the bed, spent caustic could be charged at a rate of 0.042 gal of total feed/1000 BTU. Stack gases were shown to contain less than 0.1 ppm of SO_2, no CO when the O_2 content was in excess of 8% and only 2.2 volume % of CO when the oxygen content was 5%. The effluent water contained suspended and dissolved solids and COD, but cyclone separator efficiency improvement and scrubber water recirculation show promise in improving the effluent quality. Operating costs were reported as 4.7¢/gallon of waste, or .095¢/Bbl of crude oil processed. (Lowry-Texas)

17a. Descriptors

*Oily wastes, *Incineration, Oxidation, Temperature, Evaporation, Thermal insulation, Air pollution, Water quality control, Chemical oxygen demand, Nozzles, Sludge, Particle size

17b. Identifiers

*Fluidized bed, *Caustic, Suspended solids, Loading rate

Copy available from GPO Sup Doc ... as ..., $1.50 ; ... from NTIS as PB ..., ... Price ... PB ..., $3.00 in paper copy, $0.65 in microfiche

17c. COWRR Field & Group 05D

18. Availability

19. Security Class (Report)

20. Security Class. (Page)

21. No. of Pages

22. Price

Send To: WATER RESOURCES SCIENTIFIC INFORMATION CENTER U.S. DEPARTMENT OF THE INTERIOR WASHINGTON, D. C. 20240

Abstractor Mark V. Lowry | Institution Texas Univ, Austin

WRSIC 102 (REV. JUNE 1971)

GPO 913-281

FLUID BED INCINERATION OF PETROLEUM REFINERY WASTES

by

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Mandan Refinery
Box 549
Mandan, North Dakota 58554

for the

WATER QUALITY OFFICE

ENVIRONMENTAL PROTECTION AGENCY

Project #12050 EKT
Grant #WPRD 215-01-68
March 1971

For sale by the Superintendent of Documents, U.S. G[illegible] Printing Office
Washington, D.C. 20402 - Price
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ABSTRACT

The applicability of the fluid bed incineration process for the disposal of petroleum refinery generated spent caustic and oily sludge in a commercial scale unit has been demonstrated.

Operating problems have been studied. Design and operating procedural changes are suggested.

The major process limitation stems from the loss of bed fluidity due to high particle size growth rate. Particle size growth rate is directly proportional to the particle diameter and rate of dissolved solid material charged and inversely proportional to the mass of material in the bed. The average particle diameter can be controlled by (1) collecting and continuously returning fine material to the bed, (2) utilizing an effective attriting system, and (3) limiting superficial space velocity to avoid elutriation of fines.

This report was submitted in fulfillment of Project #12050 EKT under the partial sponsorship of the Water Quality Office, Environmental Protection Agency.

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CONCLUSIONS

1. The fluid bed incineration process has been demonstrated to be practical and effective for the disposal of petroleum refinery generated spent caustic and oily sludge.

2. The process creates no atmospheric pollution problems, emitting only carbon dioxide nitrogen and water vapor. The odor of the off gas has been described by various observers as being slight to non-existent.

3. The ash produced contains sodium carbonate, sodium sulfate, other soluble inorganic salts and inert material such as sand, clay, rust, etc.

4. Operations are frequently prematurely terminated by the loss of bed fluidity caused by excessive bed particle size. The particle size growth rate can be controlled by including salient design features and by exercising proper operating techniques.

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RECOMMENDATIONS

The Mandan Refinery fluid bed incinerator is functional. Run lengths are limited by the loss of bed fluidity resulting from excessive average particle size and poor particle size distribution in the bed. The design of future units should include the following considerations:

a. Superficial Space Velocity

1. Avoid excessive elutriation.

2. Provide sufficient air capacity for fluidization of bed when starting up cold.

b. Dust Recovery

1. An efficient dust recovery system that continuously returns fines to the bed is necessary to provide small nucleii on which to deposit the chemicals thus avoiding an excessive particle growth rate.

2. Bed material lost through the dust recovery system is scrubbed out of the off gasses with water. The waste water from the scrubber is contaminated with suspended and dissolved solids from the unrecovered fines.

c. Attrition or Fines Recirculation System

An efficient easily controlled attrition or fines recirculation system must be provided so that an uncontrolled particle growth rate can be avoided.

d. Instrumentation

Continuous recording pressure differential instruments should be provided across the total fluid bed and across part of the fluid bed so that operators can control bed depth and observe bed density

e. Operation

Timely knowledge of the relative fluidity (fluid bed density/apparent bulk density) is essential if appropriate action is to be taken to avoid the complete loss of fluidity.

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INTRODUCTION

Among the many waste disposal problems incident to petroleum refinery operations, the American Oil Refinery at Mandan, North Dakota, found the problem of disposing of oily sludge and spent caustic most perplexing.

Oily sludge is characterized by the presence of oil, solids, and water. Its consistency varies from that of a fluid slurry to a greasy semi-solid. It is encountered when cleaning oil storage vessels, removed from the waste water oil-water separator and it is collected as a residue from oil-water emulsion breaking operations.

The most common disposition of this material had been to dump it into open earthen sludge pits. In order to conserve land, the volume of sludge was reduced by burning the hydrocarbon that eventually migrated to the surface of the pit. Burning produced a billowing black cloud of smoke. This method of disposing of sludge was most unsatisfactory; it degraded the land and created an obvious source of atmospheric pollution.

Spent caustic results from the use of solutions of caustic soda to treat petroleum products for the removal of acidic impurities such as naphthenic acid, cresols, mercaptans, and hydrogen sulfide. These compounds emit intense disagreeable odors. When an attempt was made to dispose of spent caustic by neutralization and by trickling the neutralized waste into the waste water treating system, it was found that the odor released during neutralization was intolerable and the resulting neutralized waste overtaxed the bio-oxidation pond, thus degrading the quality of the waste water effluent from the refinery.

When the spent caustic was dumped into an earthen pit similar to those used for the disposal of oily sludge, the caustic waste leached through the earth and created a potential source of contamination of the natural ground water. When the pit was lined with plastic sheets to prevent this migration, it rapidly became full. The volume of spent caustic was increased by rainfall. The only reduction of the waste occurred from evaporation which also produced an undesirable odor in the atmosphere.

In the long term effort to find a solution to these two waste disposal problems, various incineration systems were considered. The oxidation of spent caustic alone would require an economically prohibitive quantity of supplemental fuel and it would be difficult to prevent atmospheric pollution. The incineration of oily sludge appeared more feasible. However, when oily sludge is subjected to high temperatures, the oil evaporates. Should the resulting hydrocarbon vapors fail to ignite, they could create an extremely hazardous explosive mixture with combustion air.

Conditions offered by the fluidized bed incinerator: extremely high heat transfer efficiency, excellent mixing, and stable combustion conditions made the system most promising. Furthermore, a bed of inert fluidized solids acting as a heat sink reduced the hazard created by a mixture of unburned hydrocarbon vapor and air.

A fluidized bed incineration process had been successfully applied to the disposal of paper mill sulfite liquors. The process based on research conducted at the Battelle Memorial Institute in Columbus, Ohio, and developed by the Copeland Process Corporation of Oak Brook, Illinois, appeared to be adaptable to the disposal of petroleum refinery generated spent caustic and oily sludge.

The American Oil Company made application to the Environmental Protection Agency, Water Quality Office (EPA), for a Research and Development Grant for this work as provided in the Clean Water Restoration Act of 1966.

The American Oil Company on June 5, 1968, accepted an EPA Research and Development Grant, WPRD 215-01-68, of $170,265 to assist the estimated $354,530 project. Project objectives were to demonstrate the use of a fluid bed incinerator for the disposal of a wide variety of petroleum refinery sludges and spent caustic solutions while minimizing water and air pollution, to establish operating limitations and to determine what changes may be necessary to assure satisfactory operations.

The Mandan Refinery fluid bed incinerator was a project of the Manufacturing Department of the American Oil Company. As in the case of all new units, responsibility for the project was divided into two phases. Phase I -- engineering, construction and preparation for operation was the responsibility of the Planning and Engineering Department at Whiting, Indiana. Phase II -- operation, maintenance, post construction studies and reports were the responsibility of Mandan Refinery personnel.

This report was prepared to make the findings of the fluid bed incinerator study available in a form requested by the Water Quality Office, EPA.

DESIGN AND CONSTRUCTION

A. DESIGN

Under contract with the American Oil Company, the Copeland Process Corporation designed and constructed the fluid bed incinerator.

The unit was designed to satisfy two feed conditions:

1. During 9 months per year
   - 10 Bbl/day oil-water separator sludge
   - 10 Bbl/day emulsion treating sludge
   - 10 Bbl/day spent caustic

2. During 3 months per year
   - 10 Bbl/day oil-water separator sludge
   - 10 Bbl/day emulsion treating sludge
   - 10 Bbl/day spent caustic
   - 10 Bbl/day tank cleaning bottoms

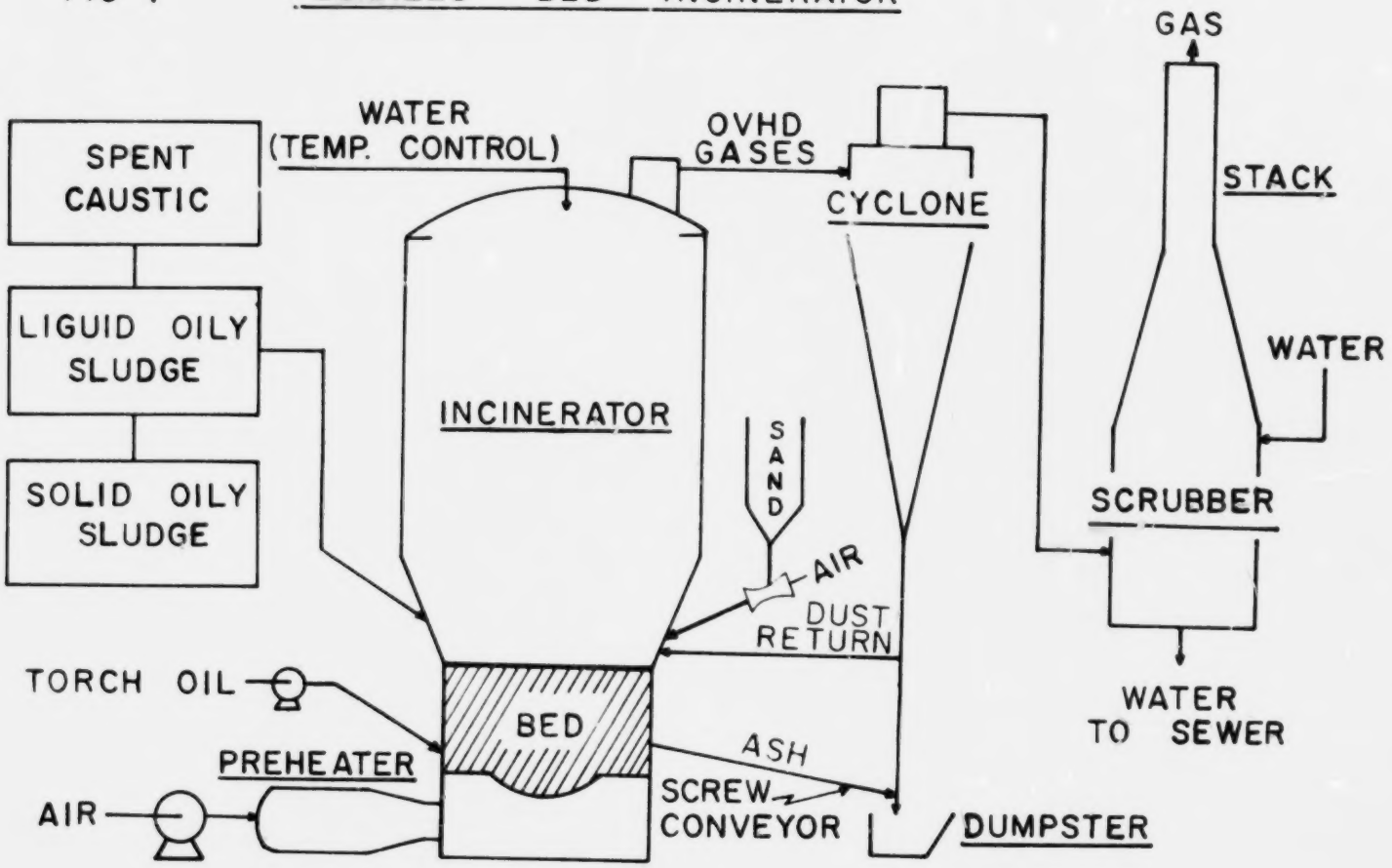
Copeland proposed that the incinerator would operate for 6 hours a day during 9 months of the year disposing of the normal production of sludge and spent caustic. During the 3-month period in which tanks are cleaned, they proposed to keep the incinerator in operation 24 hours a day.

Samples of typical sludges and spent caustic were sent to the Copeland Process Corporation and to Battelle Memorial Institute, Columbus, Ohio. The incinerator design was based on findings from examination of these samples. The Battelle Memorial Institute charged the Mandan Refinery waste material to their 22-inch diameter fluidized bed incinerator pilot plant. By these pilot plant operations, they confirmed that the spent caustic and oily sludge from the Mandan Refinery was amenable to fluidized bed incineration. However, the pilot plant exhaust gas line plugged during the test. The plugging was believed to be caused by the presence of iron sulfate. Evidently the iron was present in the sludge as rust from tank bottoms. Based on this data, Copeland designed the system to operate with a controlled reactor exhaust gas temperature to prevent scale in the exhaust duct. It was recommended that the incinerator be operated at a bed temperature of 1300 to 1500°F.

Salient design features incorporated in this unit which differ from the typical application of fluidized bed incineration to paper mill waste sulfite liquors were as follows:

1. Waste feed injection -- The feed to the waste sulfite liquor incinerator is sprayed into the vessel from the top. The spray cools the effluent flue gases as it evaporates. The dried solids fall to the fluidized bed where

FIG 1 FLUIDIZED BED INCINERATOR



they are oxidized. The oily sludge-spent caustic incinerator was designed to inject the mixed feed through a special Copeland burner nozzle just above the fluidized bed of solids. The flue gas temperature was to be controlled by injecting a cooling water spray through the top of the incinerator vessel.

2. Supplemental fuel for start-up -- The Sulfite Liquor incinerators are brought up to operating temperature by preheating combustion air with a natural gas burner. Because of the possibility of creating an explosive mixture of hydrocarbon vapor from the sludge and combustion air, the sludge-caustic incinerator design includes an additional system for injecting Torch Oil (heavy fuel oil) into the bed of fluidized solids a few inches above the bed supporting grid. When the fluid bed becomes hot enough for ignition, via the natural gas combustion air preheater, the Torch Oil is injected. It is thus possible to maintain a steady high temperature in the bed affording assurance of ideal oxidizing conditions when the sludge is introduced.

3. The inventory of material comprising the fluidized bed in the unit is normally determined by measuring the differential pressure between a point just above the grid and a point in the free board space above the bed. The Mandan incinerator design included an additional pressure tap so that by measuring the differential pressure across a portion of bed, its density could be determined. With knowledge of the bed density, the height of the fluid bed level above the grid can be determined from the differential pressure across the entire bed.

4. In order to cope with semi-solid sludges from tank cleaning, a steam heated 80 bbl. sump was included in the design. Facilities were included to stir the contents of the sump, mix it with fluid sludge from the feed tank, and to deliver the mixture to the feed tank.

Figure 1 reflects the Mandan fluid bed incinerator process flow as it was designed.

CONSTRUCTION

The Copeland Process Corporation arranged for Horton Process Division of Chicago Bridge and Iron Company to handle construction engineering (foundations, piping, and electrical), subcontractors, and field supervision.

Construction of the unit started September 3, 1968, when the Scan Construction Company, a subcontractor, began excavation for vessel foundations.

Construction was scheduled for completion February 1, 1969. However, the project was repeatedly delayed by various occurrences beyond the control of Copeland, namely:

1. A strike at Gould's Pump Company delayed the delivery of several pumps.

2. Severely cold winter season, including a 19-inch snowfall.

3. An American Oil Company Refinery turnaround. Turnaround contractors hired craftsmen away from the incinerator project.

4. The construction phase of the incinerator project was essentially completed in early June, 1969.

A photograph of the completed unit is reproduced in Figure 2.

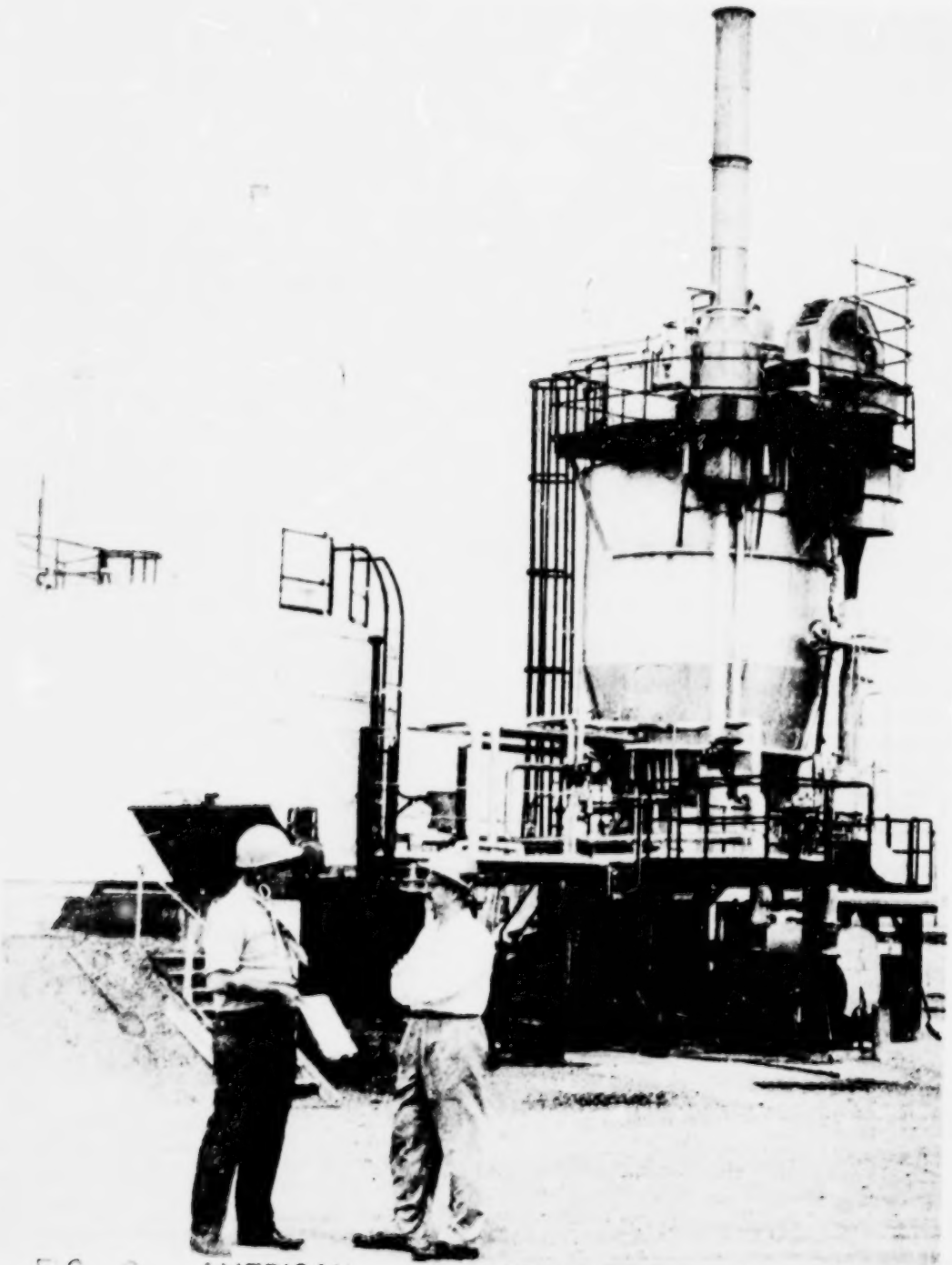


FIG 2 AMERICAN OIL COMPANY FLUID BED INCINERATOR AT MANDAN

START UP AND INITIAL OPERATIONS

Prior to starting up the fluid bed incinerator, operating instructions and operator training materials were prepared by the American Oil Company based on information furnished by Copeland. All supervisors and operators to be assigned to the unit were given 10 hours of classroom training. The Operating procedures and instructional material are attached as Appendix A.

To put the unit into operation, the air compressor is started. The air preheater is fired, and as the system is warmed, sand is introduced, forming a bed of fluidized solids. When sufficient sand has been introduced and the fluid bed is in the range of 900 to 1000°F, torch oil is admitted. The burning of torch oil causes a rapid rise in bed temperature. When the bed temperature is in the range of 1200 to 1300°F, liquid oily sludge is introduced. Provided the oily sludge contains a sufficient quantity of combustible materials, air heating and torch oil injection can be terminated. The temperature of the bed is then controlled by balancing feed rate and temperature-control water. When the operation on oily sludge has been stabilized, spent caustic is introduced.

In order for the operation to be self-sustaining, the oily sludge must contain a sufficient quantity of combustible material. About 29,000 BTU are required per gallon of total feed. The burning of sludge with a low heating value results in a lower spent caustic to sludge ratio.

As the sludge is burned, the solids remain in the bed, while the gaseous products of combustion and water vapor discharge through the gas-cleaning system. Water in the caustic solution is vaporized and the combustible material is oxidized; the solids accumulate in the fluid bed. The sodium oxide formed reacts with the products of combustion to form sodium carbonate and sodium sulfate, both of which remain in the fluid bed. The proper bed level is maintained by withdrawing ash as it accumulates from the deposition of solids. Ultimately, the original sand bed is completely displaced by the incoming solids and chemicals. The overhead gases pass through a cyclone separator from which recovered fines can be either returned to the bed or discharged along with the excess bed material. From the cyclone separator, the gases are passed through a water scrubber before being admitted to the atmosphere.

A chronological log of operations covering the period of June 20, 1969, to January 30, 1970, is attached as Appendix B.

DESIGN CHANGES BASED ON OPERATING EXPERIENCE

Wind Box Insulation

The unit was designed to conduct the hot air from the air preheater through a chamber (wind box) and air distribution grid before reaching the bottom of the fluid bed. The start-up procedure calls for heating the fluid bed to 900°F via the air preheater before injecting torch oil. During the initial attempt to start up, excessive heat loss from the wind box was observed. When the wind box temperature reached 840°F the fluid bed temperature was only 540°F. The unit was shut down and additional insulation was applied to the interior of the wind box. In subsequent start-up operations, satisfactory conditions were observed.

Air Distributor Grid Gussets

The air distributor grid was held in place by welding it to the shell of the vessel completely around its circumference. Additional support was provided by 12 equally spaced 1'-1-1/2" x 1'-1-1/2" gussets made of one-half inch steel plate. These gussets were greater in number and size than was necessary to hold the grid in place. Furthermore, they transmitted excessive heat from the grid to the shell of the incinerator vessel. When the air preheater was in operation, the exterior skin temperature of the wind box reached 850°F ultimately causing stress cracks in the vessel shell. Copeland redesigned the gussets reducing their number and size. After these revisions were completed, the maximum wind box exterior shell temperature observed while the air preheater was in operation was 240°F.

Feed Preparation

Feed handling facilities as built included a 1,000-bbl. feed mix tank, a 405-bbl. spent caustic storage tank, and an 80-bbl. tank bottoms receiving sump. It was proposed that daily pumping of sludge from emulsion breaking and from the waste water-oil separator would be received directly into the 1,000-bbl. feed mix tank. Spent caustic was to be transferred to the feed tank in batches based on the estimated oil content of the final mix.

During early operations, only oily sludge was fed to the incinerator. It was immediately apparent that the oil content (heating value) of the feed was critical. Furthermore, the oil content of sludges was observed to vary widely.

Before attempting to charge spent caustic, temporary piping was installed to permit its introduction as a separate stream. The caustic was moved by the spent caustic transfer pump and the flow rate was determined by tank gauge difference. The charge of caustic was adjusted by a hand operated valve to balance the heat released by sludge burning. With this system, it was difficult to adjust and measure the spent caustic charge but it was possible to dispose of spent caustic in this manner.

An attempt was made to operate as originally planned, by mixing caustic with sludge in the feed tank and charging a single stream to the incinerator. In order to dispose of caustic at a maximum rate, it was necessary to blend to the minimum oil content of total feed. A subsequent pumping of separator sludge that was low in oil content caused the mixed feed to fall below the minimum heating value necessary to sustain combustion in the incinerator. Feed to the unit was discontinued and the feed tank mixer was turned off. Some "free water" settled to the bottom of the feed tank and was drained to the sewer thus increasing the concentration of oil in the feed mix. While the move alleviated the problem created by having a low oil contant feed, spent caustic drained to the sewer along with the "free water".

From this experience it was concluded that spent caustic should be fed to the incinerator as a separate stream so that the caustic feed rate could be adjusted to compensate for the varying oil content of the sludge.

The Mandan incinerator is currently operating with the temporary caustic feed system. Future caustic and sludge incinerator designs should provide systems to feed caustic and sludge separately.

Oily Sludge Injection

Based on fluid bed incinerator operating experience in the paper industry, Copeland designed the unit so that sludge feed would be injected just above the fluid bed. When oily sludge was fed to the unit at this location, the oil in the sludge immediately evaporated and burned in the space above the fluid bed, thus creating a condition similar to after burning. Heat transfer to the fluid bed was poor. Furthermore, because of the high effluent gas temperature the cooling spray water flow was automatically increased, further cooling the fluid bed. While charging sludge above the fluid bed level, it was necessary to provide supplemental heat with the air preheater or by torch oil injection to maintain the proper fluid bed temperature.

When the sludge feed piping was rerouted to permit the injection of sludge through the torch oil nozzles (into the fluid bed a few inches above the grid), the bed temperature rose, cooling spray water was reduced and it was possible to maintain the proper fluid bed temperature without supplemental heat.

Spent Caustic Injection

In an effort to find the optimum point relative to the fluid bed for injecting spent caustic, four conditions were examined, namely:

1. All spent caustic was injected above the fluid bed.
2. All spent caustic was injected into the fluid bed as a separate stream.
3. Spent caustic was mixed with sludge feed for injection into the bed.
4. Spent caustic feed was split injecting part above the bed and part into the bed.

The rate at which spent caustic can be charged to the unit is limited by the heat content of the sludge. From the standpoint of maximizing the spent caustic disposal rate, there appeared to be an advantage to injecting the caustic above the fluid bed. Spent caustic was charged at a rate producing 0.042 gal. of total feed per 1,000 BTU when injected above the bed. A rate of only 0.039 gal. of total feed per 1,000 BTU was attainable while injecting caustic into the bed.

However, when caustic was injected above the bed at a rate in excess of 1 bbl per hr., bed particles adhered to the interior of the flue gas exit duct and the cyclone causing a high pressure drop across the cyclone. Successful operations were sustained by injecting up to 1 bbl per hr. above the bed and the remainder of the caustic feed into the bed.

Subsequent investigation revealed that the average chemical oxygen demand of the effluent water from the flue gas scrubber dropped by about 100 ppm when the spent caustic injection was moved from the over-bed nozzle to the in-bed nozzle. This amounts to a reduction of approximately 84 lbs. of chemical oxygen demand per day in scrubber water going to the refinery waste water treating facilities.

There appeared to be no difference in unit performance between operation with a single mixed feed of oily sludge and spent caustic injection into the bed vs. separate caustic and sludge feed streams into the fluid bed. Problems associated with mixing the caustic and sludge were previously discussed.

From these observations, it is concluded that the separate injection of spent caustic into the fluid bed produces the preferred performance of the unit.

Cooling Water Spray

The temperature of gases above the fluid bed is automatically controlled by spraying water into the top of the incinerator. The spray nozzles are purchased from Spraying Systems Company, 3201 Randolph Street, Bellwood, Illinois. Initially the unit was equipped with a one-inch G1590 nozzle that produced a spray angle of 15° at 15 psi. This spray nozzle permitted unevaporated water to reach the top of the bed. The fluid bed was thus cooled necessitating the burning of supplemental fuel to maintain proper combustion temperature in the bed.

The spray nozzle was changed to one which produced a 91° spray angle. With this nozzle, the bed temperature was not adversely affected by spray water and it was possible to operate without supplemental fuel. However, when the unit was shut down, it was found that the spray nozzle was causing impingement of unevaporated water against the vessel wall. The bed particles which became wetted agglomerated and adhered to the vessel wall in large deposits.

The 91° spray angle nozzle was replaced by a 30° spray nozzle. While this spray nozzle presented an acceptable compromise (the effluent gases were effectively cooled without necessitating the use of supplemental fuel). The ratio of caustic to sludge feed possible without supplemental fuel was lower than it had been with the 91° spray angle nozzle.

By reducing the nozzle size from one-half inch to three-eighths inch and increasing the angle to 120°, the most satisfactory results were obtained. The smaller nozzle produced a fog type spray, thus preventing the impingement of unevaporated water against the vessel wall and preventing droplets of water from reaching the fluid bed.

It is possible that the problems associated with the injection of spent caustic solution above the bed might have been solved by injecting the caustic at the proper location through a fogging spray nozzle. Some incentive exists for investigating this matter further.

It is noteworthy that Copeland recommended setting the free board temperature controller at 900°F while permitting the fluid bed to approach 1500°F. In practice it was found that free board cooling by the spray water has a tendency to cool the fluid bed even under the best cooling spray conditions. The Mandan incinerator is operated with the free board temperature controller set to hold a temperature approximately the same as the fluid bed temperature.

Feed Distribution

In the Copeland design, sludge and caustic were to be injected through a single nozzle above the fluid bed and torch oil was to be injected through a header with four equally spaced nozzles into the bed. Early experience demonstrated that a single torch oil nozzle was all that was necessary or desired. However, when sludge was fed through the torch oil system into the bed, the charge rate was limited by apparent afterburning when using a single nozzle. The operation was considerably improved by using two of the torch oil nozzles for sludge injection. When using two injection nozzles off of a single header, as one nozzle begins to become plugged the flow simply increased through the other nozzle. It is obvious to operators when one of the nozzles becomes plugged because when this happens there is an immediate rise in free board temperature, an automatic increase in cooling spray water rate, and a decrease in bed temperature.

The distribution of sludge feed into the bed through several injection nozzles is obviously necessary. It would be highly desirable to arrange the feed system to insure the equal distribution of charge through each of the feed nozzles.

Scrubber Water

In the original design of the unit, water from the refinery circulating cooling water system was utilized in the venturi-type gas scrubber. Subsequently a piping change was made permitting the substitution of raw (untreated) Missouri River water for cooling water to the scrubber. This change reduced the incinerator operating cost by eliminating the water treating chemicals which were lost via the scrubber effluent. Furthermore the chemical oxygen demand (COD) and dissolved solids (DS) content are greater in circulating cooling water than in raw water thus an equivalent reduction in COD and DS was made in the scrubber effluent water.

The scrubber effluent water contributes dissolved and suspended solids to the refinery waste water treating facilities. The amount of these materials depends on operating conditions and cyclone efficiency. In the event that these materials overtax waste water treating facilities, it is suggested that they can be greatly reduced by circulating the scrubber water.

PROCESS LIMITATIONS

Fusion Temperature of the Fluid Bed Material

The incinerator bed material primarily consists of particulate matter from the sludge plus carbonate and sulfate salts formed by chemical reactions of the spent caustic and combustion gases. In the event the bed temperature is permitted to exceed the fusion temperature of the bed material the particles become sticky. They adhere to the wall of the incinerator vessel and form large agglomerated masses in the bed. Ultimately the bed loses fluidity and it becomes necessary to shut the unit down to remove the large chunks of agglomerated bed material.

The bed material fusion temperature is periodically determined in the laboratory. The test (see Appendix C) determines the temperature at which unaerated particles of bed material adhere to each other to such an extent that they will no longer flow.

The fusion temperatures as determined by this test range from 1000°F to 1700°F. It has been found that the incinerator bed temperature can be held as much as 200°F higher than the laboratory determined fusion temperature without adversely affecting the operation.

Fluid Bed Performance

The unit was designed based on a superficial space velocity through the bed of 3 ft/sec with a bed particle size distribution similar to that represented by curve A, Figure 3. It was anticipated that the particle size would tend to increase as the dissolved solids from the spent caustic were deposited on the particles in the bed. Provided a sufficient amount of fine material was introduced with the sludge to serve as new nuclei for the deposition of these chemicals as excessive bed material was withdrawn, the bed particle size distribution would remain satisfactory.

However, when it was found necessary to inject caustic and sludge into the bed instead of above the bed, the superficial space velocity increased considerably, to as high as 5 ft/sec thus making it impossible to retain the small particles contributed by the sludge long enough for them to serve as nuclei for the deposition of chemicals. As a result the bed particles become large and relatively uniform in size. A typical size distribution is represented by curve B, Figure 3.

In spite of the large mean size and poor size distribution of the bed particles, operations have been sustained for as long as 23 days. Chronological plots of sieve analyses may be found in Figures 4, 5, and 6.

The relatively large mean particle size and poor distribution in the Mandan incinerator fluid bed has produced adverse effects on operations as follows:

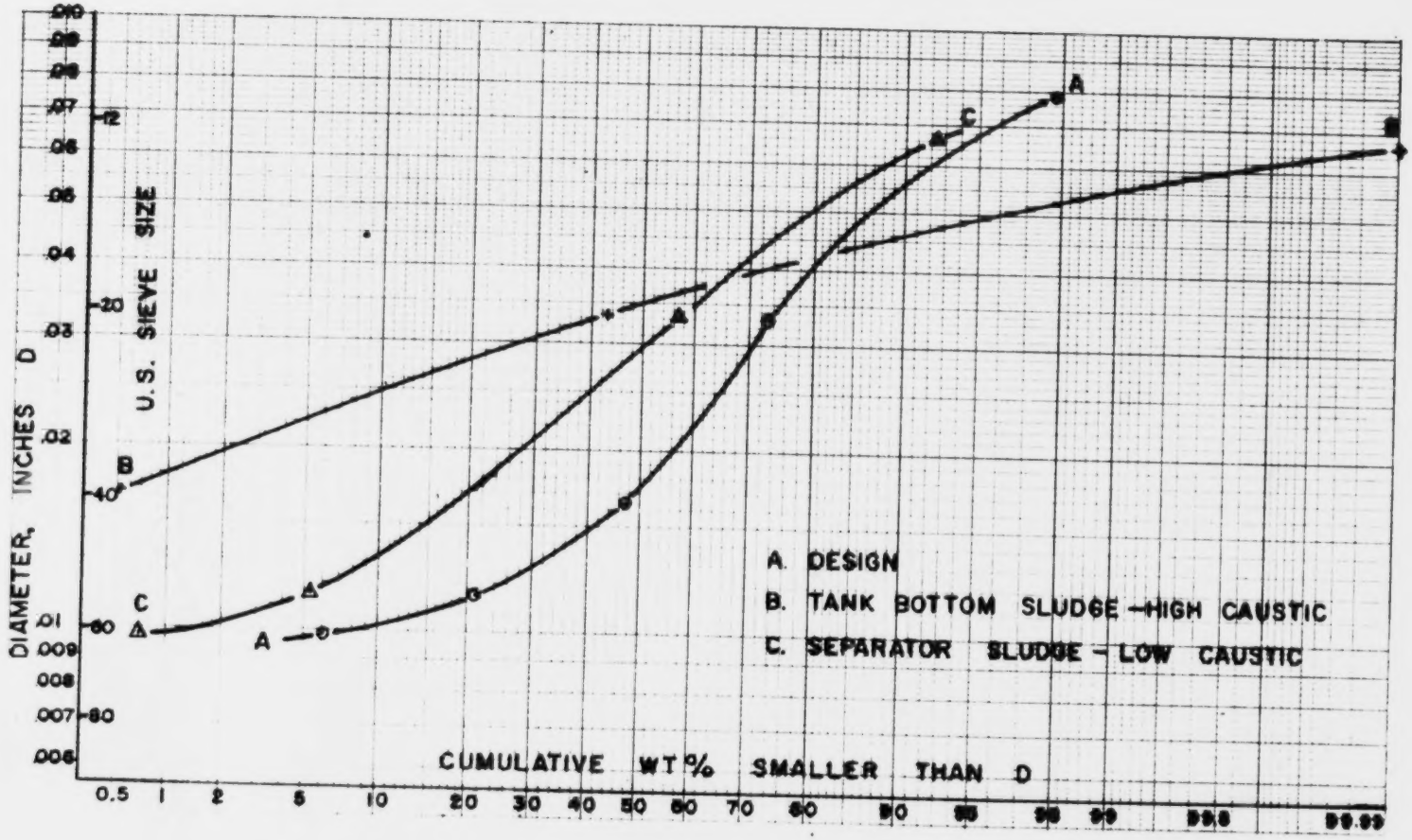


FIG 3 BED PARTICLE SIZE DISTRIBUTION

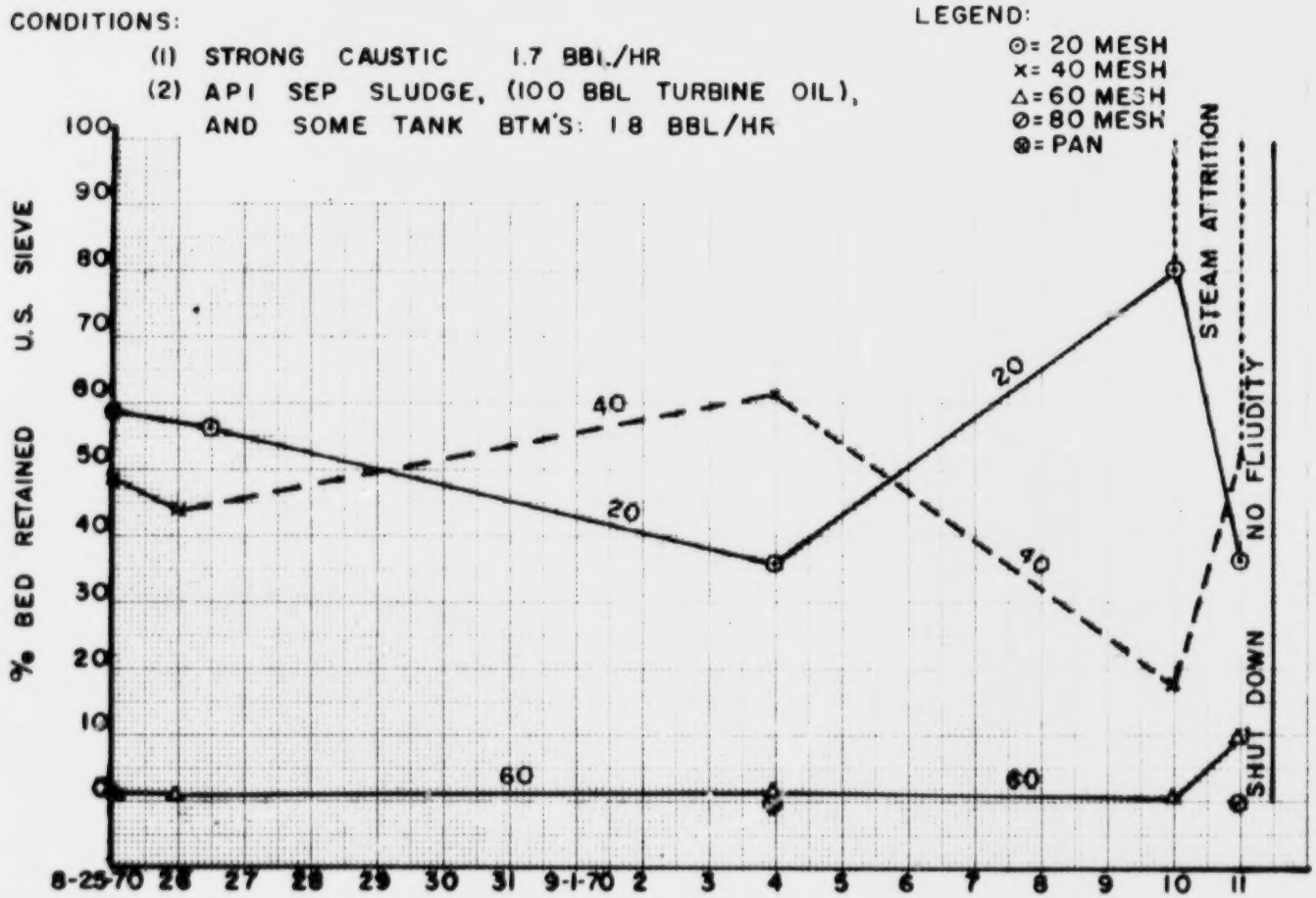


FIG 4 CHRONOLOGICAL PLOT OF BED SIEVE ANALYSIS

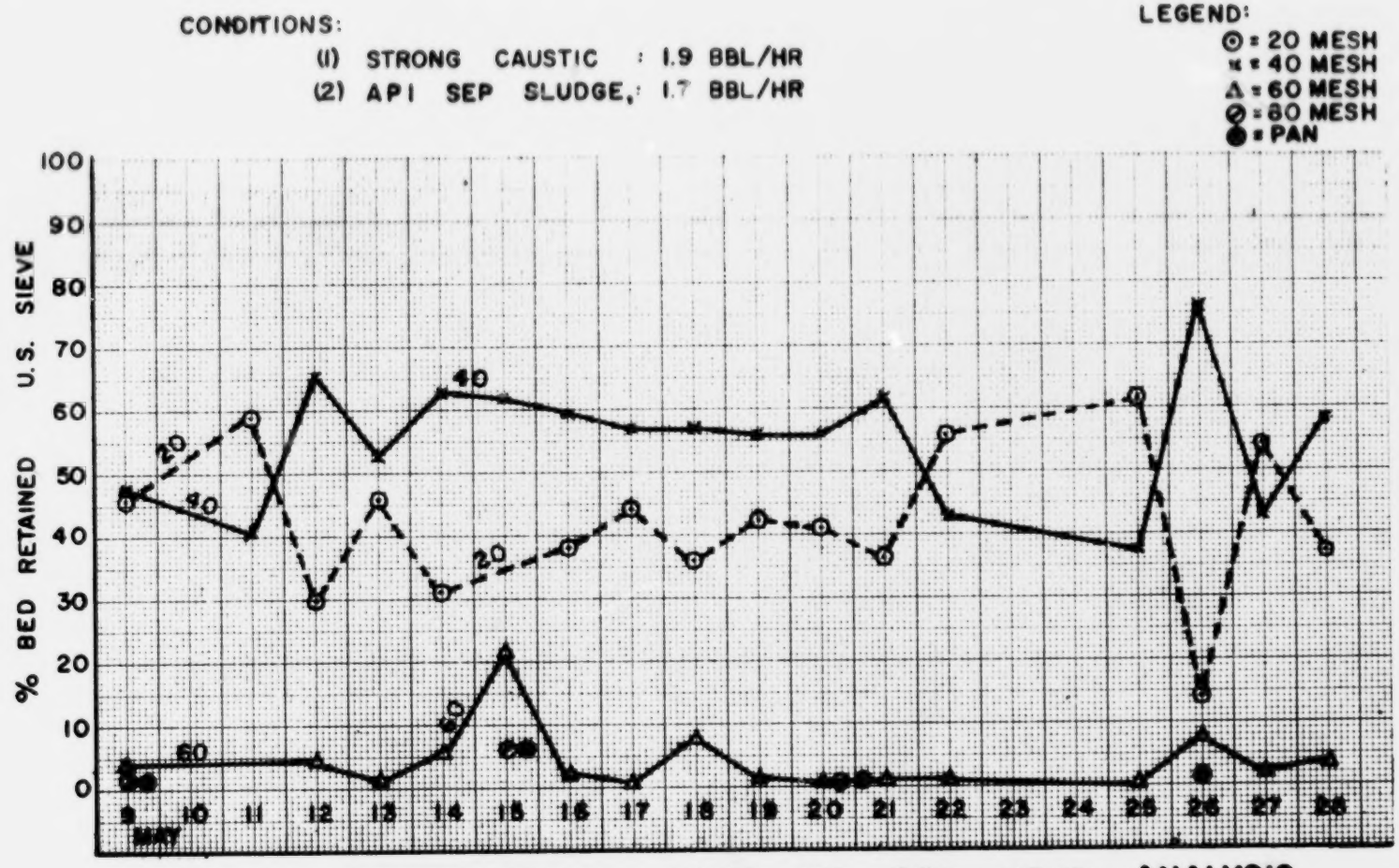


FIG 5 CHRONOLOGICAL PLOT OF BED SIEVE ANALYSIS

CONDITIONS:

(1) STRONG CAUSTIC: 0.9 BBL/HR
(2) API. SEP SLUDGE: 3.0 BBL/HR

LEGEND:

O = 20 MESH
x = 40 MESH
Δ = 60 MESH
● = 80 MESH
(P = PAN

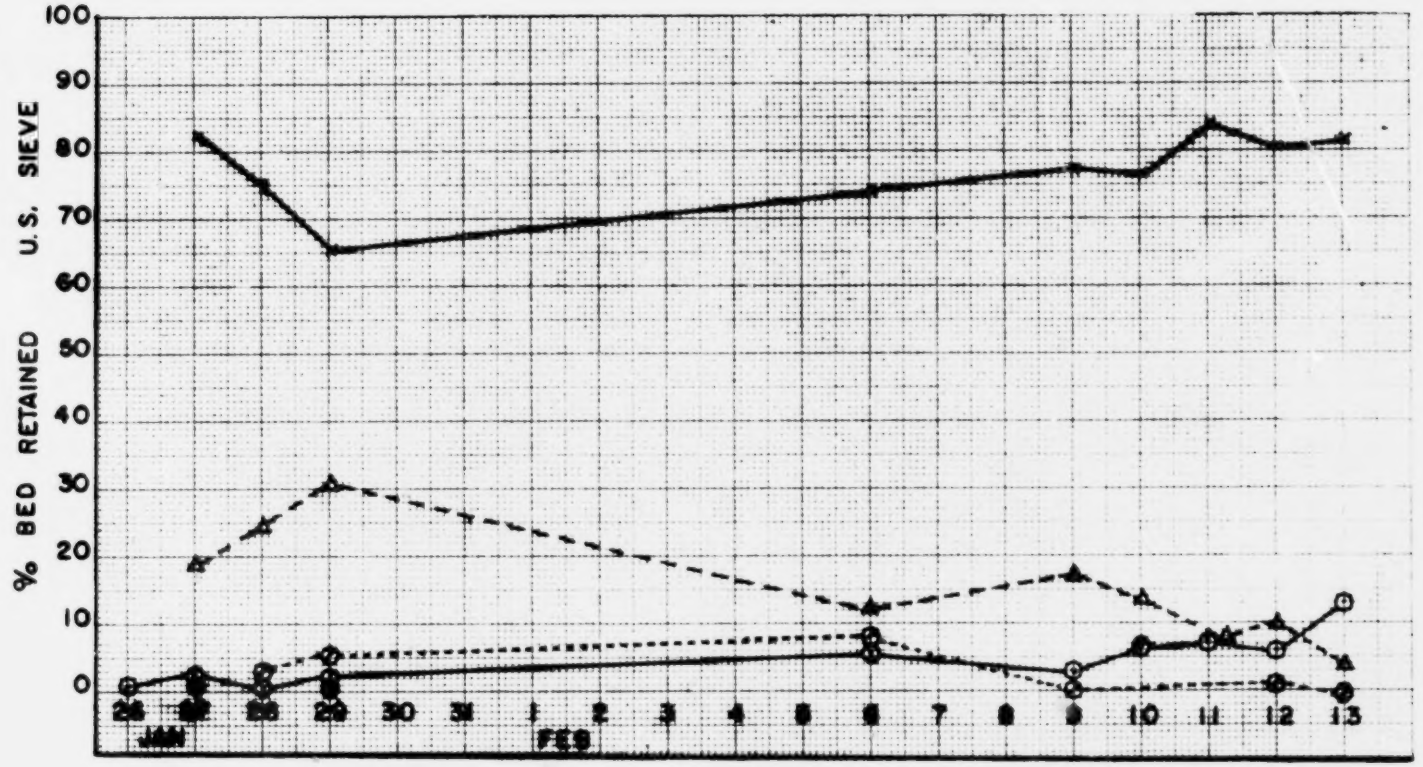


FIG 6 CHRONOLOGICAL PLOT OF BED SIEVE ANALYSIS

Thumping

The Mandan incinerator vessel emits a violent thumping sound. Shock waves are transmitted to the unit foundations. The vibrations are obvious to anyone standing near the vessel. The violence of the thumping increases as the pressure drop across the fluid bed increases. It is speculated that the bed is slugging in contrast to a relatively steady fluidized state.

Initial Fluidization of the Bed During Startup

During a startup after the fluid bed had been permitted to slump and cool, it was not possible to bring the bed to a fluidized state with combustion air. At the time it was suggested that the bed had become moist causing the particles to adhere to each other. The bed was partially removed and replaced with sand and the startup continued.

Bed material with a size distribution represented by Curve B, Figure 3, a particle density of 176 lb/ft^3, and an apparent bulk density of 88 lb/ft^3 produced satisfactory operation of the unit with a superficial space velocity of 4.74 ft/sec. The fluidized bed density was 64.4 lb/ft^3. Utilizing the smoothed correlation of particulate fluidization curves of Zenz, it was determined that the incipient fluidization velocity for this material would have been 2.03 ft/sec. Combustion air at 100°F would have produced a superficial space velocity of only 1.12 ft/sec. It would have been necessary to bring the combustion air to 565°F before the incipient fluidization velocity would have been reached. While the presence of moisture has not been eliminated as a factor contributing to the startup difficulties described above, it is highly probable that the bed simply required a higher velocity for fluidization than was possible with cool combustion air.

Torch Oil Ignition

The bed temperature at which torch oil ignites has been unpredictable. Early considerations led to prescribing a minimum bed temperature of 900°F to be reached before injecting torch oil. With the initial sand fill, this temperature proved to be adequate. However, during subsequent startups when the unit contained a chemical bed from the previous run, torch oil would not ignite at 900°F. With a 900°F fluid bed temperature and over 10 percent oxygen in the flue gas, the injection of torch oil would produce a billowing yellow cloud of smoke from the stack with no effect on bed temperature. It was necessary to raise the bed temperature to 1050°F before the torch oil would ignite.

This problem has been attributed to poor fluidization of the bed as a result of a large particle size with inadequate space velocity for proper fluidization. Apparently large bubble-like slugs of air and vaporized torch oil pass through the bed. Ignition is inhibited by poor heat transfer from the bed particles.

When the average particle size is reduced by making up part of the startup bed with fine sand, the torch oil ignites without difficulty at 900°F.

Bed Particle Size Growth

When spent caustic is injected into the incinerator, the water evaporates, combustible materials are oxidized, the alkaline materials react with the acidic gases from combustion, and the solids thus produced are deposited on the surface of the fluid bed particles causing them to increase in size. Bed particle size growth may be offset by the introduction of smaller particles along with the sludge and by natural attrition occurring in the bed. However, in spite of these offsetting actions, often the bed particle size has been observed to increase quite rapidly ultimately causing a complete loss of bed fluidity. Attempts to reduce the particle size by jetting steam into the bed has met with limited success.

Particle growth rate is related to the mass of the bed, the rate of mass addition, and the particle diameter as follows:

$$R = \frac{W\,D}{3M} \quad \text{(See Appendix D for proof)}$$

R = rate of increase in particle diameter, in/hr
W = rate of mass addition to the bed, lb/hr
M = mass of bed material, lb
D = diameter of bed particles, in

Since the rate of particle growth is directly proportional to the diameter of the particles in the bed, the average particle size must be controlled to keep the growth rate at or below the rate of particle size reduction by attrition and fines additions.

When attempts are made to reduce the average particle size by injecting steam into the bed, there is an apparent improvement. However, because of the loss of fines through an inefficient cyclone with an inadequate system for returning fines and a less-than adequate method of steam attrition, the average particle size in the bed cannot be reduced enough to halt the spiraling effect of particle diameter on growth rate and fluidity is ultimately lost.

Loss of Fluidization

The interruption of operations is most frequently caused by the loss of bed fluidity. At the outset, it was recognized that the introduction of spent caustic would cause some particle size growth and some changes in the bed characteristics. However, it was believed that particles introduced with the sludge would serve as new nuclei for the deposition of chemicals and as excessive bed material was withdrawn a tolerable equilibrium would be reached.

Instrumentation for following bed conditions included a temperature recorder on six thermocouples located around the bed at various depths and a bed differential pressure indicator. In addition a piping arrangement was provided to permit switching the upper pressure differential tap to a lower point so that periodically the differential pressure across 30 inches of fluid bed could be measured.

At 3 a.m., November 21, 1969, the incinerator lost fluidity and it was necessary to shut the unit down. Fluid bed data from the unit log for 42 hours preceding the shutdown are plotted against time in Figure 7. Observing only the total bed differential pressure and temperature spread, an obvious indication of trouble was displayed by the temperature spread among the thermocouples in the bed. One can ordinarily expect bed temperature variations of not more than $\pm$ $10^{o}F$. At about noon on November 19, 1969, a temperature difference between the hottest and coldest thermocouple was recorded at $67^{o}F$. Fluidizing air was increased from 1,290 SCFM to 1,600 SCFM and the bed temperature spread returned to normal. The temperature spread jumped again at 6 p.m. Spent caustic was discontinued for about 45 minutes with no immediate apparent effect. Steam was jetted into the bed at about 7:45 a.m. and again the temperature spread in the bed returned to normal. The attrition steam was shut off at 3 a.m. on November 20, 1970. The temperature spread in the bed increased but the temperature differences were not alarmingly high. However, the bed temperature spread continued to increase and steam was reintroduced at 11 a.m. on November 20, 1969. This time attrition steam had little apparent effect. Finally, at about 3 a.m. on November 21, 1969, the bed temperature spread jumped to $1000^{o}F$. The pressure drop across the bed fell. The unit was shut down.

Fortunately, throughout this period regular recordings were made of the pressure differential across 30 inches of the bed. The partial bed differential (Δp) is related to fluidized bed density ϱb in lb/ft^3 as follows: (see Appendix E)

$$\varrho b = 2.079 \ \Delta p$$

NOTE:
SUCCESSIVE OBSERVATIONS ARE CONNECTED BY STRAIGHT LINES FOR CONVENIENCE AND DO NOT IMPLY KNOWLEDGE OF INTERMEDATE VALUES.

LEGEND
X = BED DEPTH
Δ = BED TEMP. SPREAD
+ = TOTAL BED DIFF.
⊙ = BED DENSITY

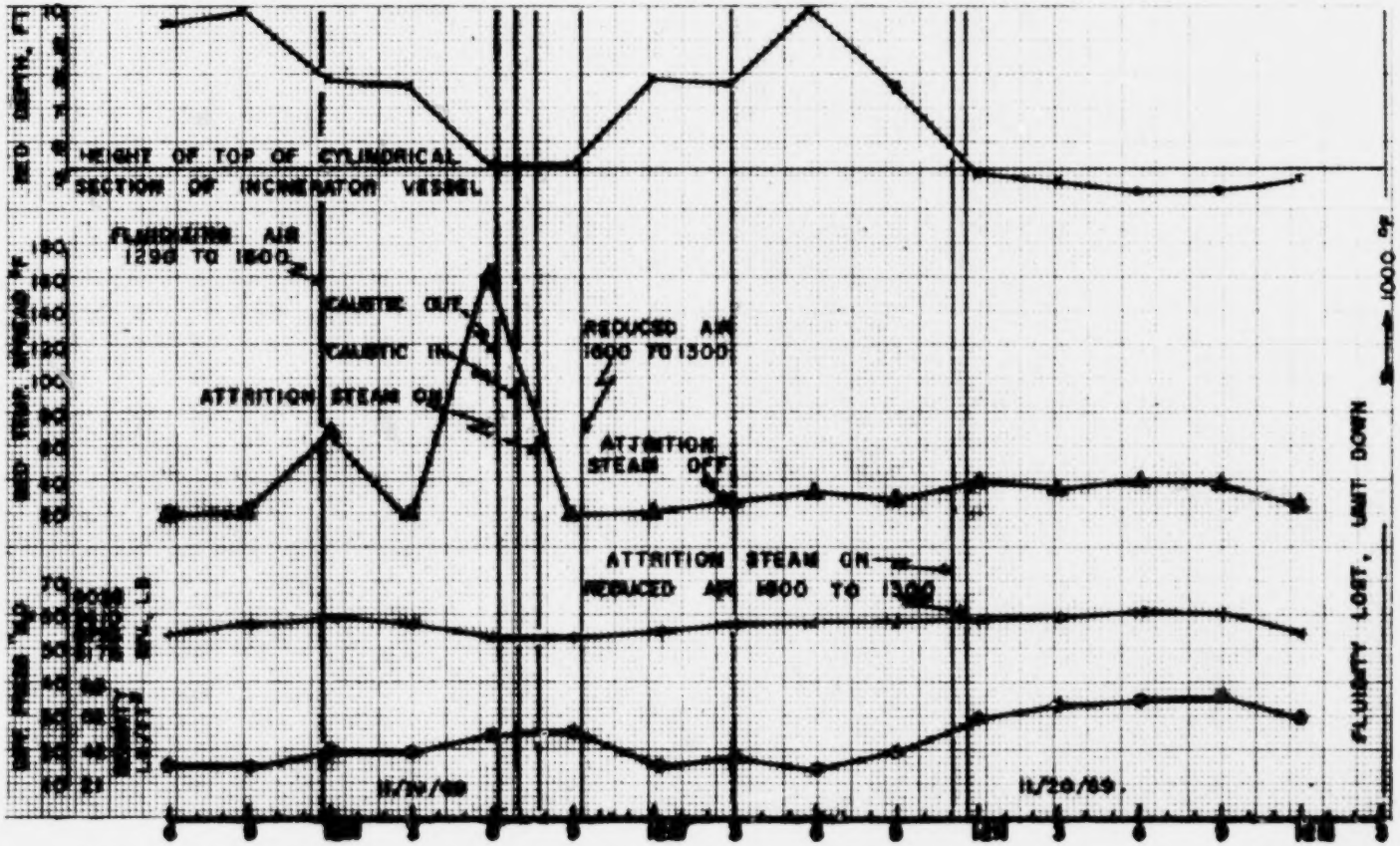


FIG 7 OPERATING DATA ON FLUID BED 11/19/69 TO 11/21/69

It will be noted that as the pressure drops across the fluid bed the superficial space velocity increases and expands the bed. Therefore one might expect the bed to be denser at the bottom. Some error is introduced when ρ_b is used to represent the average density of the entire bed.

The total depth of the bed (H) in feet may be determined from the total bed differential pressure (Δp) and the partial bed differential (Δp) as follows: (see Appendix E)

$$H = 2.5 \frac{\Delta P}{\Delta \overline{P}} + 0.5$$

NOTE: The lower differential pressure tap is 0.5 ft. above the grid.

Inventory of bed material (M) lb. above the lower differential tap may be determined from the total bed differential.

$$M = 123.55 \ \Delta P \text{ (see Appendix E)}$$

This equation is valid so long as the entire bed is contained in the lower cylindrical portion of the bed. When the bed extends up into the frustum of the expanded section of the vessel, an obvious error is introduced due to the shape change.

Bed density and bed level data are also included in Figure 7. A relatively wide variation in bed density is observed. Bed density is reduced by an increase in superficial space velocity (increased air) but without changes in space velocity the fluidized bed density displayed an overall tendency to increase. The bed depth is certainly not predictable from the total bed differential alone but its inverse relationship to bed density is apparent. Note that on two occasions during this period, the bed level reached a height of over 10 feet. The top of the cylindrical section of the incinerator vessel is at 5.5 feet. When the bed level is over 5.5 feet, the top of the bed extends into the inverted frustum of the expansion section of the vessel. The thumping previously described, in addition to being related to the bed particle size, bed height, and diameter, is probably influenced by the superficial space velocity change that occurs in the frustum.

The additional knowledge gained by observing the fluid bed density is essential to the determination of the depth of the fluid bed. However, the relative state of bed fluidity is not apparent from any combination of the data in Figure 7. Note that the first indication of poor fluidity shown by an increased bed temperature spread occurred at about 12 noon on November 19, 1969, at a time when the bed density was 52 lb/ft^3. The bed density reached a maximum at 73 lb/ft^3 between 6 and 9 p.m. on November 20, 1969, and was apparently declining when the fluidity was lost.

When a bed of finely divided solids is in a fluid state, the fluid bed density decreases as the superficial space velocity is increased and conversely as the superficial space velocity is decreased, the fluid bed density increases. The maximum bed density that can be reached by reducing velocity is represented by the density of the mass of particles at rest. This is called the apparent bulk density (ABD). Since fluidization causes an expansion of the mass of particles, the ratio of fluid density to apparent bulk density serves as a measure of relative fluidity. Since fluidity was nearly lost according to the data in Figure 7, at various fluid densities it follows that the apparent bulk density of the bed must have been changing.

Apparent bulk density is a function of the particle packing configuration and in a system containing uniformly sized particles, it is independent of particle size. In a system containing variously sized particles, the apparent bulk density decreases as uniformity increases the limiting value being the same whether the uniformity approaches the large or the smaller of the sizes in the distribution. For example, a mass of uniformly sized spheres of diameter D arranged so that each sphere is in contact with six other spheres would have an ABD expressed by:

$$ABD = 0.5236\ \rho_p \text{ (see Appendix F)}$$

Where ρ_p is the particle density

If the spheres are arranged so that each sphere is in contact with 12 other spheres:

$$ABD = 0.700\ \rho_p$$

Note that in neither case is the ABD related to the size of the spheres.

When the packing configuration is such that each sphere is in contact with six other spheres, small spheres whose diameter is 0.4142 D can be added to the mass without increasing the volume until the void space is filled. As the smaller diameter spheres are added, the ABD of the mixture will increase from $0.5236\ \rho_p$ to a maximum then decrease to $0.5236\ \rho_p$. The calculated ABD of mixtures of spheres with diameters 0.064 and 0.0265 and having a particle density of 150 lb/ft^3 are recorded in Table I.

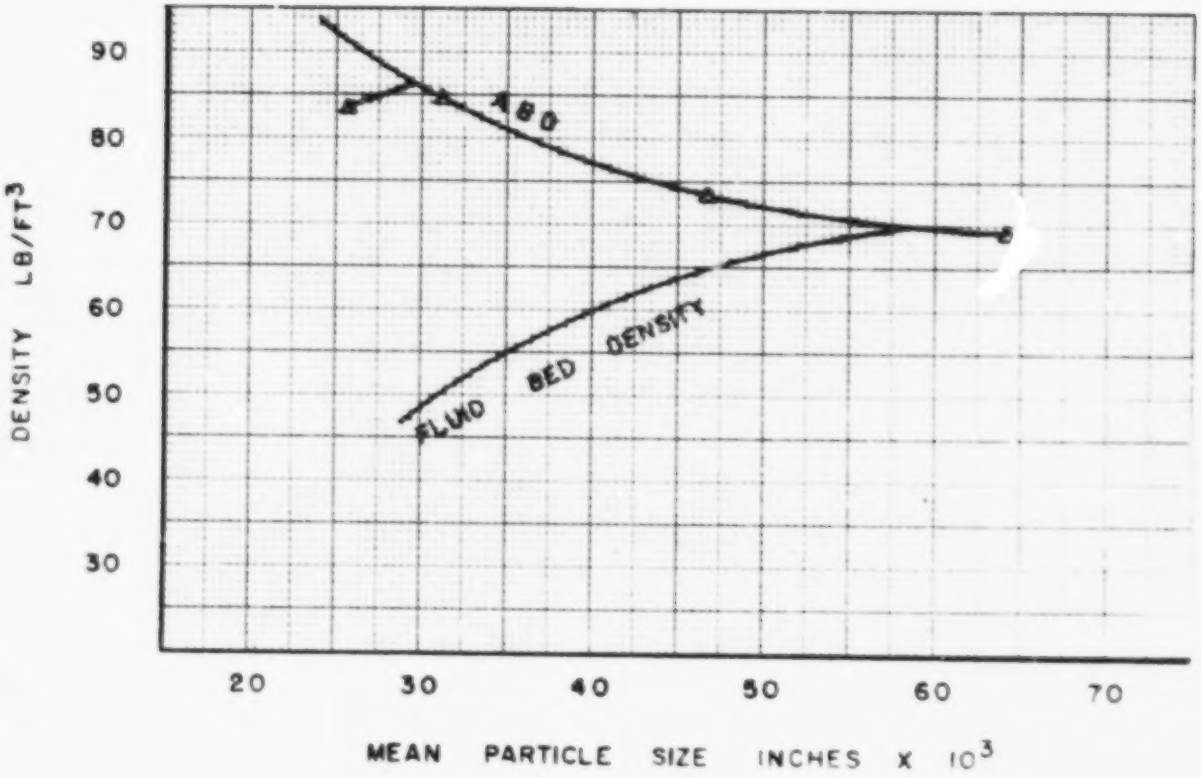


FIG 8 AVERAGE PARTICLE SIZE VS ABD AND FLUID BED DENSITY

TABLE I

CALCULATED ABD OF MIXTURES OF SPHERES 0.064 IN. & 0.0265 IN. DIAMETER WITH PARTICLE DENSITY OF 150 LB/FT^3

% 0.064 D	% 0.0265 D	Calculated ABD lb/ft^3	Calculated Weight Mean D, in.	Observed ABD lb/ft^3
100	0	78	0.064	70
81	19	96	0.056	-
68	32	115 max.	0.052	-
63	37	112	0.050	-
58	42	108	0.048	-
47	53	100	0.044	74
13	82	85	0.033	-
5	95	80	0.031	85
0	100	78	0.026	84

When a mass of finely divided particles are poured into a container, it would be highly unlikely for them to come to rest in a definite configuration such as was assumed to find the calculated density. These calculations serve only to provide an understanding of variations observed in ABD with various mean particle sizes. Several observations are included in Table I.

The density of a fluidized bed is related to the superficial space velocity, the mean particle size, the particle density and the fluidizing medium density and viscosity. Fluidized bed densities have been estimated for various average particle sizes utilizing the smoothed correlation of particulate fluidization curves. Variables arbitrarily held constant were as follows:

Fluidizing media density	$\rho_f = 0.0212$ lb/ft^3
Particle density	$\rho_p = 150$ lb/ft^3
Viscosity of media	$\mu_f = 2.82 \times 10^{-5}$ lb/ft-sec
Space velocity	$V_1 = 4.21$ ft/sec

Estimated fluid bed densities are tabulated in Table II.

TABLE II

Estimated Fluid Bed Densities
For Various Mean Particle Sizes

Particle Diameter, in. D_p	Fluid Bed Density lb/ft^3
0.029	46.9
0.032	50.7
0.036	56.6
0.040	60.0
0.044	61.6
0.048	64.9
0.052	66.7
0.056	70.2

Observed ABD from Table I and fluid bed density from Figure 2 are plotted vs. mean particle size in Figure 8. When the average particle size increases as the result of the loss of fine material the ABD decreases and the fluid bed density increases. When the fluid bed density equals the ABD the bed ceases to be fluid. For an operation in which the particle size is increasing due to the deposits of solids, fine material is being lost and new fines material is being added to the system, it is impossible to predict the ABD.

It is proposed that untimely termination of operations can be avoided by appropriate action if instrumentation is provided to furnish a continuous record of fluid bed density and the relative fluidity is determined periodically by comparing fluid bed density with ABD. (ABD can be simply measured by weighing a known volume of bed material and converting the volume weight relationship to lb/ft^3.)

The ability to control the ratio of the rate of fine material additions to the rate of dissolved solids additions is essential. See Appendix H.

TABLE III
MANDAN FLUID BED INCINERATOR
PERFORMANCE DATA

	TEST	I	II	III	IV	V	VI
	DATE	10/29/69	11/19/69	3/12/70	3/17/70	3/23/70	5/25/70
FEED							
Sludge							
Oil, wt %		34.0	36.6	29.3	31.6	22.1	22.2
Water, wt %		61.1	50.8	56.7	56.4	62.0	30.8
Sediment, wt %		4.9	12.6	14.0	12.0	15.9	47.0
Sulfur, wt %		0.67	0.57	0.78	1.25	1.98	0.56
Sed. Ignition Loss, wt %		50.0	46.5	21.3	13.8	7.8	24.2
Sed. Ash, wt %		50.0	53.5	78.7	86.2	92.2	75.8
Sp. Gr. Oil		0.8723	0.8763	0.8850	0.8840	0.8880	0.9290
Heating Valve Sludge, BTU/gal.		47,191	56,108	42,054	42,553	43,671	60,848
Spent Caustic							
Water, wt %		68.8	81.0	65.5	62.4	72.2	71.6
Alkalinity) Hydroxide NaOH, wt %		8.9	8.8	8.4	8.2	7.4	6.1
Alkalinity) Carbonate Na_2CO_3 wt %		7.7	7.6	10.3	10.2	9.5	7.7
Alkalinity) B. Carbonate $NaHCO_3$, wt %		0	0	0	0	0	0
Total Sodium, wt %		5.8	1.5	8.6	7.8	7.1	1.3
Total Sulfur, wt %		2.7	2.6	2.4	3.4	2.9	2.9
COD, Mg/l		110,000	73,600	76,400	146,000	124,000	197,500

TABLE IV
MANDAN FLUID BED INCINERATOR
PERFORMANCE DATA

	TEST	I	II	III	IV	V	VI
	DATE	10/20/69	11/19/69	3/13/70	3/17/70	3/23/70	5/25/70
BED MATERIAL							
Bed							
U.S. Sieve. % Ret.	(20, wt %	25.4	82.2	16.0	38.2	5.1	57.0
	(40, wt %	73.2	11.8	77.3	60.5	80.3	42.5
	(60, wt %	1.4	0	6.0	1.3	14.6	0.5
	(80, wt %	T	0	0.7	0	0	0
	Pan, wt %	T	0	0	0	0	0
	Water Soluble, wt %	49.2	75.3	65.0	76.6	77.3	64.4
	Water Insoluble, wt %	50.8	27.4	35.0	23.4	22.7	35.6
	Sulfate $SO_4^=$, wt %	37.1	41.1	41.2	41.2	46.6	39.3
	Sodium Na^+, wt %	15.4	23.2	14.7	22.0	21.6	20.0
Alkalinity	(Hydroxide NaOH, wt %	0.8	0	0.2	1.0	0.3	1.4
	(Carbonate Na_2CO_3, wt %	2.0	10.2	2.5	4.9	2.4	4.0
	(B. Carbonate $NaHCO_3$ wt %	0	T	0	0	0	0
Cyclone Fines							
U.S. Sieve. % Ret.	(20, wt %	24.2	7.4	50.4	37.0	7.6	55.8
	(40, wt %	43.5	17.8	18.8	30.8	34.8	12.3
	(60, wt %	5.3	3.1	3.6	5.2	8.1	4.9
	(80, wt %	3.6	3.1	2.4	3.6	5.1	3.6
	Pan, wt %	23.4	68.6	24.8	23.4	44.4	23.4
	Water Soluble, wt %	28.0	77.3	49.8	63.3	72.8	-
	Water Insoluble, wt %	72.0	22.7	50.2	36.7	27.2	-
	Sulfate $SO_4^=$, wt %	23.1	32.6	32.0	38.9	39.3	-
	Sodium Na^+, wt %	7.4	7.2	-	16.4	20.4	-
Alkalinity	(Hydroxide NaOH, wt %	0	0.9	0	0.6	1.0	-
	(Carbonate Na_2CO_3, wt %	0.8	12.2	2.3	3.2	2.7	-
	(Bicarbonate $NaHCO_3$, wt %	0.3	0	0.8	0	0	-

TABLE V
MANDAN FLUID BED INCINERATOR
PERFORMANCE DATA

	TEST	I	II	III	IV	V	VI
	DATE	10/20/69	11/19/69	3/13/70	3/17/70	3/23/70	5/25/70
SCRUBBER							
Off Gas							
ORSAT	CO Vol %	2.2	0	0	0	0	0
	CO_2 Vol %	10.8	4.4	5.4	6.2	4.4	7.0
	O_2 Vol %	5.0	8.0	11.0	10.0	15.4	11.0
Hydrocarbon, PPM Vol		-	-	0	0	0	0
SO_2, PPM Vol		-	-	0	T	T	T
Particulates Grams/SCF (Dry)		-	-	0.058	0.040	0.046	0.098
Influent Water							
pH		6.9	6.9	7.0	6.7	6.4	7.8
Suspended Solids		24.0	30.0	36.0	18.0	26.0	58.0
Dissolved Solids		764.0	938.0	1108.0	1054.0	934.0	1148.0
COD		72.0	64.0	56.0	168.0	52.0	124.0
Effluent Water							
pH		6.0	6.0	6.9	6.6	6.5	7.5
Suspended Solids		722.0	530.0	321.0	207.0	283.0	2406.0
Dissolved Solids		1169.0	1628.0	2092.0	1394.0	1564.0	1792.0
COD		144.0	160.0	28.0	60.0	36.0	312.0

TABLE VI
MANDAN FLUID BED INCINERATOR
PERFORMANCE DATA

OPERATING CONDITIONS	TEST DATE	I 10/20/69	II 11/19/69	III 3/13/70	IV 3/17/70	V 3/23/70	VI 5/25/70
Average Bed Temp, ^{o}F		1302	1308	1274	1286	1271	1318
Temperature Deviation $\pm$ ^{o}F		5	16	6	8	8	2
Average Freeboard Temp, ^{o}F		1296	1353	1381	1381	1374	1346
Average Freeboard Pressure, in H_2O		13	24	21	18	20	15
Average Bed Differential, in H_2O		60	56	71	66	66	73
Average Density 30" Bed, in H_2O		18	20	-	-	-	31
Average Fluid Air, SCFM		1342	1356	1416	1372	1357	1463
Sludge Rate, bbl/hr		I 2.9	I 2.5	I 1.9	I 1.9	I 1.8)	I 2.1
Caustic Rate, bbl/hr		A 0.6	A 1.4	A 1.3	I 1.1	I 0.7) 2.5	I 2.0

A = Above Bed
I = In Bed

TABLE VII
MANDAN FLUID BED INCINERATOR
MISCELLANEOUS INFORMATION

TEST	I	II	III	IV	V	VI
DATE	10/20/69	11/19/69	3/13/70	3/17/70	3/23/70	5/25/70
BTU/Gal Total Feed	39,101	35,966	24,969	26,952	31,443	31,165
Sludge lb/gal	8.28	8.94	9.21	8.89	9.48	15.52
Spent Caustic lb/gal	9.88	9.86	10.08	10.05	9.91	9.07
Barometric Pressure "Hg	29.98	30.25	30.35	29.91	29.74	29.99
Space Velocity, ft/sec Air at grid	2.73	2.69	2.68	2.69	2.64	2.89
Feed Injec.	3.46	3.37	3.22	3.20	3.18	3.73
Top of Bed	2.15	2.00	-	-	-	3.82
Freeboard	1.49	1.45	1.44	1.43	1.45	1.55
Fluidized Bed Density, lb/ft^3	37.4	41.5	-	-	-	64.5
Bed Depths, ft	8.3	8.1	-	-	-	5.9

UNIT PERFORMANCE

Test Data

In order to evaluate the performance of the fluid bed incinerator, six sets of test data were observed during operations. These data are recorded in Tables III through VII. Sludge feed rates were observed between 1.8 and 2.9 bbl/hr and caustic rates varied between 0.7 and 2.0 bbl/hr. Heat requirements per gallon of total feed ranged from 25,000 to 39,000.

The ash produced (bed material) is relatively inert. The maximum hydrate alkalinity observed was 1.4 wt %. Solubility ranges between 50 and 80 wt % (in hot 212^{o}F water). The possibilities of utilizing this material or reclaiming chemicals from it exist but have not been investigated.

Stack gases contain water vapor, carbon dioxide, oxygen, and nitrogen. Only traces of sulfur dioxide (less than 0.1 ppm) were found. On one occasion, 2.2 volume % carbon monoxide was detected when the oxygen content was 5 percent. CO was not evident when the O_2 content was over 8 percent.

The scrubber effluent water contains suspended solids, dissolved solids, and chemical oxygen demand. These waste water contaminants would be greatly reduced by improving the cyclone separator efficiency and recirculating scrubber water.

Heat and Material Balance

Heat and material balance around the incinerator vessel were calculated for Tests IV and VI. These data are summarized in Table VIII. Note that in Test IV the heat released per gallon of sludge charged (44,272 BTU/gal) by heat balance is in the same order of magnitude as the estimated heating value of the sludge (42,509 BTU/gal) but the (42,509 BTU/gal) heat release by heat balance in Test VI does not agree with that estimated from the sludge composition (60,848 BTU/gal).

In estimating the heating value of sludge, the percent oil is determined and its heating value is estimated from the quality of oil recovered. The sediment is ignited and the volatile matter from the sediment is assumed to be oil. The sediment in the sludge for Test IV was only 12 wt % with 12.8 percent ignition loss while that in Test VI was 47 wt % with 24.2 ignition loss. The assumption that the material lost from the sediment by ignition has the same heating value as the oil was obviously not valid for Test VI. A more accurate value of sludge heating value could be established by use of a bomb calorimeter.

TABLE VIII
MANDAN FLUID BED INCINERATOR
HEAT AND MATERIAL BALANCE

		TEST IV					TEST VI				
		Lb/Hr	T, °F t -60°F	Cp BTU/ Lb/°F	H BTU/Lb	BTU	Lb/Hr	T, °F t -60°F	Cp BTU/ Lb/°F	H BTU/Lb	BTU
Sludge 120°F	Oil	235.92	60	0.420		5,945	459.58	60	0.420		11,581
	Water	400.12	60		59.82	23,935	421.61	60		59.82	25,221
	Ash	73.38	60	0.224		986	487.67	60	0.224		6,554
Caustic 90°F	Water	289.73	30		) 29.93	8,672	545.52	30		) 29.93	16,327
	NaOH	38.07	30		)	1,139	46.47	30		)	1,390
	Na_2CO_3	47.36	30	0.256		365	58.66	30	0.256		451
	DS	89.15	30	0.2		535	111.23	30	0.2		667
Air (180°F)	O_2	1,785.60	120	0.223		47,782	1,888.00	120	0.223		50,522
	N_2	5,920.89	120	0.249		174,674	6,259.55	120	0.249		187,035
Spray (76°F)	Water	600.00	15		14.97	8,982	334.90	15		14.97	5,013
		9,480.19				273,014	10,613.19				304,761
Off Gas (1381°F IV) (1346°F VI)	CO_2	683.32	1321	0.2802		252,927	833.36	1286	0.280		300,076
	O_2	801.60	1321	0.253		267,904	952.64	1286	0.253		309,949
	N_2	5,878.04	1321	0.274		2,127,580	6,214.60	1286	0.274		2,188,391
	H_2O	1,829.11	1321		568.1	1,039,117	1,859.94	1286		564.4	1,049,750
Ash (1286°F IV) (1318°F VI)		73.38	1226	0.224		20,152	487.67	1258	0.224		137,421
	DS	89.15	1226	0.200		21,859	111.23	1258	0.200		27,985
	Na_2SO_4	109.55	1226	0.202		27,130	132.22	1258	0.202		33,599
	Na_2CO_3	16.04	1226	0.256		5,034	21.53	1258	0.256		6,933
		9,480.19				3,761,703	10,613.19				4,054,104
Heat Release by Difference						3,488,689					3,749,343
BTU/Gal Sludge from Heat Balance						44,272					42,509
BTU/Gal Estimated from Sludge Composition						42,553					60,848

OPERATING AND MAINTENANCE COSTS

Automatic controls are provided to allow the unit to operate with minimum attendance by an operator, and to shut down the unit in the event of a hazardous malfunction. Under normal operations, approximately 25% of an operator's time is required for this unit. When tank bottoms are being transported to the unit, the number of man hours required may increase appreciably.

Utilities required for the unit appear reasonable. The connected electric load is approximately 48.2 horsepower per ton of sludge feed, most of which is required for the air blower. Natural gas and torch oil are required for startup only. Compressed air (100 psig) at 13,000 SCFH, steam (140 psig) at 900#/Hr, and raw water usage at approximately 35,000 #/Hr are the only other utilities required.

Installation of piping to facilitate using raw water in place of treated cooling water has reduced water costs approximately 70%.

Maintenance costs will average approximately 1.7 per cent per year of installed cost per year. The two major contributors to maintenance cost are feed nozzle plugging and pump wear caused by the abrasiveness of the sludge.

The following is a breakdown of projected operating costs based on a yearly operating factor of 33.2%: The expenditure pattern in Table IX provides additional cost information.

Operating	Maintenance	Utilities	Total
$4,813	$5,854*	$6,653**	$17,320

A yearly operating cost of $17,320, processing 8,712 Bbl/Year of waste caustic and sludge, results in a unit cost of 4.7¢/gallon of waste. The Mandan Refinery has a crude running capacity of 50,000 Bbl/Day. Based on crude running capacity, this cost is 0.095¢ per Bbl of crude processed. It is noteworthy however, that the Mandan Refinery processes a low sulfur (0.4 Wt.%) 42° API gravity crude containing little residual fuels or asphaltic material. The production of spent caustic and oily sludge is therefore at a minimum compared with the refining industry.

*Based on average expenditures for quarter ending June 30, 1970. These initial costs are high for this prototype unit, and will undoubtedly decrease in the future.

**Water costs are based on using untreated water.

TABLE IX
MANDAN INCINERATOR
EXPENDITURE PATTERN

A. Construction

	Quarter	Construction (Copeland + American)	Engineering	Total
1968				
May-June		(Included on 3rd Quarter Expenditures)		
July-Sept.	3	$ 61,123	$ 8,428	$ 69,551
Oct.-Dec.	4	96,829	15,341	112,170
Total 1968		$157,952	$23,769	$181,721
1969				
Jan.-March	1	$ 87,950	$(-2,045)[a]	$ 85,905
April-June	2	30,510	1,943	32,453
July-Sept.	3	29,491	1,305	30,796
Oct.-Dec.	4	2,466	1,126	3,592
Total 1969		$150,417	$ 2,329	$152,746
1970				
Jan.-March	1	$ 660	$ 0	$ 660
April-June	2	182	0	182
Total		$ 842	$ 0	$ 842
Grand Total		$309,211	$26,098	$335,309

(a) Correction for inadvertent additional engineering charge for 4th quarter of 1968.

B. Operation

Months	Quarter	Operation & Postconstruction Maintenance	Studies & Reports	Utilities	Total
1969					
April-June	2	$ 731	$ -	$ -	$ 731
July-Sept.	3	4,824	3,257	1,795	9,876
Oct.-Dec.	4	7,683	3,848	2,345	13,326
Total		$13,238	$ 7,105	$ 4,140	$23,933
1970					
Jan.-March	1	$ 4,317	$ 5,180	$ 3,015	$11,292
April-June	2	7,647	5,180	4,113	16,203
Total		$11,964	$10,360	$ 7,128	$27,495
Grand Total		$25,202	$17,465	$11,268	$53,935

ACKNOWLEDGEMENTS

Dr. Roy N. Giles, former Manager of the American Oil Refinery at Mandan, retired, insisted that a system for disposal of refinery generated oil sludges and spent caustic must be found.

K. C. Lowe, Project Manager, American Oil General Engineering, was responsible for administration of all aspects of engineering and construction.

W. E. Meihack, Engineer, American Oil Mandan Refinery Engineering and Technical Division was responsible for site preparation and design of outside battery limits facilities. He provided liaison between the American Oil Company and construction contractors.

C. H. Ruzicka, General Foreman American Oil Refinery at Mandan, represented the refinery Operating Division during the planning and construction stages and supervised start up operation.

H. P. Jasper, Operating Foreman American Oil Refinery at Mandan prepared the operators training material and initial operating procedures.

J. H. W. Haig and T. R. Green, Operating Foremen American Oil Refinery at Mandan supervised the initial operations of the fluid bed incinerator. Their reported observation provided the basis for much of this report.

Olev Kraav, Chief Chemist, American Oil Refinery Engineering and Technical Division developed analytical procedures and supervised laboratory work associated with the fluid bed incinerator.

Cecil M. Wheeler, Vice President Copeland Systems Incorporated provided constructive comments on portions of this report during its preparation.

R. F. Reiche, Engineer and Representative of Walker Process Equipment, a Division of Chicago Bridge and Iron Company, arranged for permission to reproduce copies of the original process flow and instrumentation drawing.

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GLOSSARY

API SEPARATOR: An oil-water separator for removing oil from waste water the design of which has been approved by the American Petroleum Institute.

APPARENT BULK DENSITY (ABD): The weight per unit volume relationship of a mass of finely divided solids at rest.

ATTRITION: The breaking up of bed particles to form smaller particles (occurs by thermal shock and/or impingement).

ELUTRIATION: The flushing of the smaller particle from the bed by the flushing medium.

FLUID BED: A mass of finely divided solids suspended in a rising stream of fluid.

FLUID BED DENSITY (ϱb): The weight per unit volume relationship of a mass of finely divided solids in a fluidized bed.

FUSION TEMPERATURE: The minimum temperature at which finely divided solids begin to stick together (fuse) on contact with each other.

INCIPIENT FLUIDIZATION VELOCITY: The minimum superficial space velocity necessary to change a mass of finely divided solids from a slumped condition to a fluidized condition.

PARTICLE DENSITY (ϱp): The weight per unit volume relationship of an individual particle.

SLUMP: The at rest condition of a fluid bed having an insufficient quantity of fluid rising through it to suspend the solid particles.

SUPERFICIAL SPACE VELOCITY: A calculated rate ($\frac{\text{distance}}{\text{time}}$) determined from the volume of fluid passing through a cross-sectional area of bed in a unit of time neglecting the space occupied by the finely divided solids.

TORCH OIL: Any heavy petroleum hydrocarbon mixture suitable for burning in a fluidized bed for temperature control.

APPENDIX A

OPERATING PROCEDURES AND INSTRUCTIONAL MATERIAL

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FLUIDIZED BED INCINERATOR

I. PROCESS DESCRIPTION

A. General

The technique of fluidized bed processing has been used in refinery cat crackers for many years. Its application to the waste disposal of paper mill effluents was developed by the Copeland Process Corporation, and they have applied the same principle to the disposal of oil refinery waste sludges. The refinery oily sludges have enough heating value to sustain combustion without the addition of supplementary fuel; as a matter of fact, provisions have been made to add water to the system to control operating temperatures.

The fluidized bed media is a mixture of properly graded granuals of a variety of materials including sand and/or chemicals. If the waste to be burned contains no inorganic chemicals the bed may be sand. However, if chemicals are present, they will be collected and agglomerated, and actually become the bed material after some period of operation. A buildup of chemical bed material will require periodic removal to maintain level of fluidity.

The operating temperature of the bed will be dependent upon the fusion characteristics of the bed material, consistent with complete combustion. If the bed material were to remain sand, the temperature may be over 1750°F.; however, low fusion chemical beds may require temperatures around 1300°F. to prevent defluidization due to fusion.

B. Application to Mandan Refinery

The incinerator for the Mandan Refinery will dispose of wastes from four sources:

API Separator Box Sludge
Emulsion Sludge from Separator Slop Tanks
Spent Caustic
Tank Cleaning Bottoms

These streams will contribute a mixture of oil, water, solids, and dissolved chemicals in various proportions.

I. PROCESS DESCRIPTION (Cont'd)

B. Application to Mandan Refinery (Cont'd)

In order to provide a feed stream of relatively constant thermal value, the wastes need to be blended and kept mixed in the feed tank. The spent caustic is stored in a separate tank at the incinerator site and will be pumped into the feed tank as required for blending and disposal. The tank cleaning bottoms will be hauled to the incinerator site by truck and dumped into a sump box where they will be heated and slurried to a pumpable mixture and then pumped into the feed tank as required for blending. The sludges from the separator area will be periodically pumped directly into the feed tank and will be blended off with the proper amounts of caustic and bottoms sludge. The rate of sludge feed to the incinerator vessel will depend upon the oil content of the feed - the higher the oil content, the slower the feed rate. During periods of extensive tank cleaning programs, the incinerator will operate for extended periods of time because of the high oil content of the tank bottoms material. During the remainder of the year the incinerator will operate until the feed tank is emptied and will stay shut down until the feed tank becomes full again. It will require approximately 5 days to consume a "batch" of feed. The laboratory will analyze the contents of the feed tank for oil content to help obtain the desired blend of feed for the incinerator.

C. Incinerator

The sludge is pumped into the incinerator and enters immediately above the expanded fluidized bed. At this point the oils in the feed are vaporized and ignited by the burning material in the bed, and solids in the feed enter the bed and are burned there. Steam and "over-bed" combustion air are introduced concentrically around the feed nozzle and serve to atomize the feed for better combustion. The oil content of the feed is purposely blended to provide sufficient heat to sustain combustion. Water sprays in the top of the vessel will cool effluent gases to a point where entrained solids will not be soft and sticky. These sprays are automatically adjusted by a temperature-recorder controller.

Air for combustion enters the vessel in two locations. A fairly constant supply of air is admitted below the

I. PROCESS DESCRIPTION (Cont'd)

C. Incinerator (Cont'd)

distribution grid and is required to keep the bed "fluidized." Sufficient sludge feed must be maintained into the incinerator to consume this "fluidizing air". A second supply of air is admitted above the bed concentrically with the feed nozzle. This air can be varied as required to provide proper combustion for increased feed rates or higher oil content feeds. Both of these air streams are supplied by electrically driven centrifugal air blower and each stream is adjusted by a flow-indicator controller. The air requirements are monitored by an oxygen analyzer sampling the effluent gases.

During start-up of the incinerator a gas fired air preheater is used in the fluidizing air stream to heat the bed to 900°F. At this point a torch oil (decanted oil) is injected into the fluidized bed to increase the bed temperature to an operating temperature of approximately 1350°F. When the operating temperature is reached sludge feed can be started and the torch oil and gas preheater shut off.

The effluent gases carrying entrained solids pass from the incinerator into a cyclone where almost 100% of the entrained particles over 20 microns in diameter will be removed. These particles removed by the cyclone can be either hauled from the unit as solid waste or returned to the incinerator bed to aid in particle size control.

The effluent gases from the cyclone are water washed in a scrubber which will remove 98% of all particles from zero to 20 microns in diameter. The cyclone and scrubber are both fabricated from 18-8 stainless steel.

Initially the incinerator will use sand as the bed material. It is expected that a static bed of four feet will expand to approximately six feet when fluidized. The "orifice plate" or "grid" contains 117-1/4" sch. 40 stainless steel pipes 16" long with a 4" long 1/2" diameter "tee" welded at the top of each. These "tees" provide for the distribution of the fluidizing air and prevent the bed material from dropping through the grid when not fluidized. After the introduction of spent caustic in the feed the sand will gradually be displaced by oxidized sodium salts of sodium

I. PROCESS DESCRIPTION (Cont'd.)

C. Incinerator (Cont'd.)

carbonate and/or sodium sulfate. The formation of sodium sulfate will depend upon the amount of sulfur present in the sludge feed. The bed will continue to grow in depth as sludge containing caustic is burned and will have to be removed periodically via the water cooled screw conveyor.

As the bed is converted to chemicals its physical characteristics will change, which will affect its fusion point and fluidizing properties. The fusion point of various mixtures of sodium compounds is between 1300 - 1350°F. Some of the iron salts known to be present in the sludge feed will further alter the fusion temperature. Periodic laboratory tests of the bed material will establish the fusion temperature and help to determine the operating temperature of the incinerator bed. Fusion of the bed will cause defluidization and subsequent shutdown for clean out. The approach of defluidization is indicated by a loss in differential pressure across the bed.

Proper fluidization is also dependent upon a proper distribution of particle sizes in the bed media. Most of the particles should range from 50 to 20 mesh (which is .012" to .033" in diameter, about the size of particles found in coarse sand). A sand bed maintains its particle distribution quite well and requires only monthly additions of sand to maintain level. However, when chemicals are present in the feed, the small sand particles tend to grow as they become coated with the chemical salt until all particles achieve the same size. At this point fluidization becomes unstable and the bed may collapse. To help maintain a better distribution of the fine particles and also to provide a nucleus for agglomeration of these salts on a continual basis, provision has been made to educt the fine particles separated in the cyclone back into the incinerator bed. It may also be necessary to dump a portion of the bed via the unloading screw conveyor and add batches of sand at some undetermined frequency to maintain proper particle size distribution.

II. INSTRUMENT CONTROL SYSTEMS

A. Storage Tanks

The "spent caustic" storage tank and the "feed" tank conventional gauge boards for level indicators. Both tanks have steam heating coils in them with automatic temperature controls to maintain the contents at 100°F. The 405 bbl. caustic tank is insulated. The 1000 bbl. feed tank is equipped with a side entry 20 HP mixer.

The "tank bottoms" sump is an 80 bbl. steel box 10 feet square and four feet deep. The cover is hinged on one side to permit dumping material into it from a dump truck. The contents are to be maintained at 160°F. by manually controlling steam to the steam grid beneath the floor. The level is indicated by means of a bubbler-type level indicator. The sump is equipped with a heavy-duty top entering mixer to mix the contents into a pumpable slurry.

B. Sludge Feed Pump

The sludge feed pump is a Moyno pump equipped with a variable speed drive. This is a positive displacement pump. The pumping rate is controlled from the Ultraformer control room by means of an air manual loading station (HIC -40) controlling the variable speed drive. The feed pump motor is shut down automatically by a temperature switch (TE-50M and TS-50H) if the incinerator freeboard temperature exceeds 1400° or drops below 1000°F. (The upper portion above the bed is the "freeboard" area.) The motor trips out on loss of fluidizing air pressure.

C. Incinerator

The sludge feed is blended to be slightly rich in oil content to sustain combustion without the need for supplementary fuel. Water must be added to control the temperature of the effluent gases so that entrained solids will not be soft or sticky to cause plugging of the cyclone. Temperature loop 49 (TE-49, I/P-49, TRC-49 and TCV-49) will automatically control the freeboard temperature at approximately 1100°F. by spraying cooling water into the upper area of the incinerator.

II. INSTRUMENT CONTROL SYSTEMS (Cont'd)

C. Incinerator (Cont'd)

In case of excessive bed temperature, emergency water will be injected into the sludge feed line. This control loop (TE-90, TS-90, and SOV-90) will be automatically activated if the bed temperature reaches 1450°F.

Both high and low temperature alarms in the freeboard zone (TE-50M, TS-50L, TA-50L, TA-50H) will warn of impending feed pump shutdowns by alarming at 1350°F and 1050°F respectively, which is 50°F before TS-50 shuts down the feed pump.

A 12 point temperature recorder will monitor temperatures in the incinerator and associated equipment.

The air preheater is designed to heat the incoming fluidized air to 1100°F so that the bed material can be heated to this temperature during starting up. Fuel for the air preheater is natural gas. Interlocks in the preheater control system will stop or prevent the injection of natural gas if any of the following occur:

Low fluidizing air pressure;	PS-75; 3	psi
Low natural gas pressure;	PS-73; 7	psi
High natural gas pressure;	PS-74; 20	psi
Loss of burner flame;	Flame sensing device	

Alarms and panel indications activate on loss of preheater flame, or the functioning of any of the pressure interlocks. The outlet temperature of the preheater is manually controlled by varying the gas so that the incinerator can be heated gradually on startup at approximately 100°F per hours.

Four torch oil injection nozzles are provided 90° apart and one foot above the grid. Decanted oil can be injected into the bed after the bed reaches 900°F to bring the bed up to 1350°F, at which temperature the sludge feed can be started to the unit. The operation of the torch oil injection is manually controlled except for the provision that the pump motor trips out on loss of fluidizing air pressure. A restriction orifice, RO-86, provides for circulation of oil in the torch oil lines by bypassing oil to the tank.

II. INSTRUMENT CONTROL SYSTEMS (Cont'd)

C. Incinerator (Cont'd)

The torch oil tank is steam traced and insulated and should manually be controlled at from 150 - 200°F.

The pressures in the incinerator are measured (1) below the grid, (2) one foot above the grid, (3) top of the incinerator, and (4) outlet of the cyclone. The following pressures are transmitted to the panel in the control room:

1) Differential across the grid, DPI-77
2) Differential across the bed, DPI-78
3) Differential across the cyclone, DPI-79
4) Pressure below the grid, PI-47
5) Freeboard pressure, PI-48

DPI-77 differential pressure will indicate if the tubes in the grid are plugging. DPI-78 differential pressure will give an indication of bed depth plus a measure of fluidity. DPI-79 differential pressure will indicate if the cyclone is plugging. An increase in PI-48, freeboard pressure, without a change in DPI-79, differential across the cyclone, will indicate an increased pressure drop across the scrubber.

Air for combustion is controlled by two flow indicator controllers, FIC-41, and FIC-42, which are panel mounted in the control room. Air through FIC-42 enters the incinerator below the grid via the air preheater and acts as fluidizing air for the bed media in addition to providing air for combustion. This air will be maintained at approximately 1255 SCFM under normal operating conditions. The over-bed air, controlled by FIC-41, enters the incinerator concentric with the sludge feed nozzle just above the bed level. This air will be varied to suit the feed rate and will be determined by the oxygen content in the flue gas as monitored by the oxygen analyzer O_2I-61. The overbed air should be adjusted to give 4 to 6% oxygen in the flue gas. An oxygen alarm is set to annunciate if the oxygen drops to 3%.

Additional controls are furnished on the compressor to aid in starting, protect against periods of low flow, and to prevent overloading. The compressor is electric motor driven and therefore must be unloaded while the compressor is started. This is accomplished by a time delay relay,

II. INSTRUMENT CONTROL SYSTEMS (Cont'd)

C. Incinerator (Cont'd)

TDR-80, which keeps the control valve SOV-80 in the compressor suction line closed for 15 seconds until the motor reaches its operating speed of 3600 RPM. The discharge pressure is controlled by PIC-80 which throttles control valve PCV-80 in the compressor suction. If the compressor is operated at too low a flow, it will surge and put undue stress on the bearings and internals. To prevent surging, control loop 60 (FE-60, FIC-60, and FCV-60) will automatically vent the compressor to atmosphere if the flow drops to the set minimum flow.

III. EQUIPMENT OPERATING INSTRUCTIONS

A. Spent Caustic Storage Tank 902-1

1. Operating temperature 100°F.
2. Maximum operating gage 375 bbls.
3. Minimum operating gage 60 bbls (to keep heater covered).
4. Keep tank block valve No. 20 closed except when filling.
5. Keep suction block valve No. 15 closed except when transferring.
6. Test condensate from heating coils each Saturday on the 8-4 shift with pH paper to check for leak in heating coils.
7. OSBL caustic line to tank 902-1 shall be blown with steam into tank 902-1 after each transfer into tank 902-1.

B. Feed Mix Tank 902-2

1. Operating temperature 100°F.
2. Maximum operating gage 900 bbls.
3. Minimum operating gage 230 bbls. to cover mixer prop.
4. Keep block valve No. 26 on sludge fill line closed except when pumping from API Separator.
5. Keep tank block valve No. 24 on spent caustic transfer line open. Block line in at spent caustic pump discharge valve, No. 18.
6. Keep tank block valve No. 41, from bottoms transfer pump closed except when transferring from tank bottoms sump.
7. Keep block valve No. 40 to suction of bottoms transfer pump closed except when transferring to tank bottoms sump.
8. Keep tank mixer No. 905-1 turned on at all times.
9. Keep suction block valve No. 25 closed except when sludge feed pump is in operation.
10. Do not pump Separator Slop Tank sludge into feed tank while incinerator is operating.

C. Tank Bottoms Sump 902-3

1. Operating temperature 100 to 160°F, depending upon material in sump.
2. Maximum gage --- 4 ft.
3. Minimum gage --- 1-1/2 ft., upper suction nozzle.
4. Control operating temperature by throttling steam

III. EQUIPMENT OPERATING INSTRUCTIONS (Cont'd)

C. Tank Bottoms Sump 902-3 (Cont'd)

coil inlet valve No. 47 (globe valve).
5. Use upper suction valve No. 49, as normal suction line.
6. Keep both suction valves closed except when transferring from sump.
7. Operate mixer 905-2 whenever preparing a slurry mix or transferring from the sump.
8. Keep cover closed except while unloading into the sump.
9. Do not unload into sump while transferring from sump.

D. Torch Oil Storage Tank 902-4

1. Tank will be filled with Decanted Oil via truck, barrels and hand pump.
2. Operating temperature 170° to 200°F.
3. Maximum operating gage 265 gallons.
4. Minimum operating gage -- refill at 100 gallons.
5. Control operating temperature by throttling steam coil inlet valve No. 149. Steam coil is on outside of tank.

E. Spent Caustic Transfer Pump 901-1

1. Keep pump suction and discharge valves closed when pump is not in service. Valves No. 15 and 18.
2. Pump is equipped with a mechanical seal.
3. Maintain level in oiler using 5W-20 oil.
4. This is a centrifugal pump. Start pump with suction valve wide open and discharge valve cracked open. Open discharge valve to establish flow.

F. Sludge Feed Pump 901-2

1. This is a positive displacement pump. Both suction and discharge valves and block valve at incinerator must be wide open before starting pump motor. Valves No. 25, 23, and 114.
2. A variable speed drive is used to vary the speed of the pump. Flow is proportional to pump speed. Pressure is independent of pump speed.
3. The pump speed can be varied from 61 RPM to 615 RPM which will vary the flow from 1/2 gpm to 5 gpm. The pump speed can be varied over its full range with Manual

III. EQUIPMENT OPERATING INSTRUCTIONS (Cont'd)

F. Sludge Feed Pump 901-2 (Cont'd)

Loading Station HIC-40. (See figure 4 for speed vs. capacity vs. air loading.)

4. Liquid will not flow backwards through this pump.
5. The pump must not be run dry. Liquid pumped is necessary as a lubricant and coolant.
6. Shaft packing should be only tight enough to prevent leakage.
7. The gear case lube level should be maintained with American Industrial Oil No. 21.
8. The pump motor will be shut down automatically by temperature switch TS-50, if the incinerator freeboard temperature exceeds 1400° or drops below 1000°F.
9. The pump motor will be shut down automatically by pressure switch PS-88 if fluidizing the air pressure drops below 3 psig.
10. Do not operate speed changer when pump is not running.
11. Pump motor switch must be in "auto" position during incinerator operation.

G. Bottoms Transfer Pump 901-3

1. This is a positive displacement pump. All suction and discharge valves must be lined up for flow before starting pump motor.
2. Liquid will not flow backward through pump.
3. A variable speed drive is used to vary the speed of the pump. Flow is proportional to pump speed. Pressure is independent of pump speed.
4. The pump speed can be varied from 58 RPM to 350 RPM to obtain a flow of from 3 to 19 gpm.
5. The pump must not be run dry. Liquid pumped is necessary as a lubricant and coolant.
6. Shaft packing should only be tight enough to prevent leakage.
7. The gear case lube level should be maintained with American Industrial Oil No. 31.
8. Do not operate speed changer when pump is not running.

H. Torch Oil Circulation Pump 901-4

1. This is a positive displacement gear pump. Suction and discharge valves No. 151 and 193 must be open before

III. EQUIPMENT OPERATING INSTRUCTIONS (Cont'd)

H. Torch Oil Circulation Pump 901-4 (Cont'd)

starting pump. Valves No. 133, 134, 135 and 136 to torch oil nozzles should be closed.
2. Torch oil will circulate back to tank through restriction orifice RO-86.
3. The pump motor will be shut down automatically by pressure switch PS-88 if fluidizing air pressure drops below 3 psig.
4. Pump bearings are grease cup lubricated with Stanobar Grease S.
5. A relief valve is built into the pump end-plate.
6. Decanted oil is used for torch oil.
7. Pump motor switch must be in "auto" position when torch oil is being used.

I. Feed Tank Mixer 905-1

1. Mixer should be in operation at all times.
2. Incinerator must be shut down if mixer fails during incinerator operation.
3. Operator shall check on the operation of the mixer every 3 hours when the incinerator is in operation.
4. Keep weight loaded stuffing box lubricator filled with Amobar Grease S through grease fitting.

J. Tank Bottoms Sump Mixer 905-2

1. Operate mixer whenever preparing a slurry mix or transferring from the sump.
2. Mixer impeller elevation approximately level with upper suction nozzle.

K. Air Compressor 304-1

1. Maintain compressor bearing oil level with 5W-20.
2. To start compressor:
   a. Close overbed air flow control valve FCV-41 with FIC-41 on manual.
   b. Fully open fluidizing air flow control valve FCV-42 with FIC-42 on manual.
   c. Check to see that control valve PCV-80 on compressor suction is closed. TDR-80 should cause the valve to close when the motor is not running.

III. EQUIPMENT OPERATING INSTRUCTIONS (Cont'd)

K. Air Compressor 304-1 (Cont'd)

d. Set PIC-80 for 9 psi and in the "automatic" position.
e. Set FIC-60 at 1100 scfm and place in "automatic" position. If total air flow drops below this point vent control valve FCV-60 will vent air to the atmosphere to prevent compressor from surging. Vent will be open during startup of compressor and close automatically at 1100 scfm.
f. Start compressor. Check to see that after 15 seconds suction valve PCV-80 opens and PIC-80 controls discharge pressure at 9 psig. Vent valve FCV-60 should close off automatically if a flow is established through the bed.
g. Set FIC-42 to desired flow but not below 1100 scfm which is necessary to fluidize a hot bed. Maximum flow will be necessary to fluidize a cold bed on startup.
h. When additional air is required for combustion, place overbed air FIC-41 into operation.

3. To Stop Compressor

a. Perform preliminary steps of cooling incinerator to a bed temperature of 700°F.
b. Check overbed air control valve FCV-41 closed.
c. Check that FIC-60 is set at 1100 scfm and in the "automatic" position.
d. Slowly shut off fluidizing air flow with FIC-42. This should cause FIC-60 to open the vent valve to the atmosphere.
e. Shut off the compressor. PCV-80 control valve in the suction line should go closed when the motor is tripped and the vent valve FCV-60 should open.
f. If in step "d" above the compressor should surge while shutting off fluidizing air, trip of the compressor immediately and then close fluidizing air control valve FCV-42. Check for a malfunction or improper setting of FIC-60.

L. Scrubber 501-1

1. Scrubber must be in operation before heat is added to the incinerator vessel.

III. EQUIPMENT OPERATING INSTRUCTIONS (Cont'd)

L. Scrubber 501-1 (Cont'd)

2. Close sample valves No. 95 and 98 and collection tank drain valve No. 180.
3. Open valve No. 96 to fill collection tank to overflowing, then close valve No. 96.
4. Open scrubber drain valve No. 190.
5. Open valve No. 72 to activate pressure gage PI-62.
6. Open valves No. 73, 74, and 75 to admit water to the venturi throat weir and sprays. Adjust valve No. 75 to hold 52 psi on sprays.
7. An increase in pressure on PI-62 indicates plugging of sprays.
8. Do not take scrubber out of service until air blower has been shut down.
9. To shut down scrubber close water valves 73, 74 and 75 to venturi throat and open valve No. 180 to drain collection tank.

M. Cyclone 303-1 -- Ejector 301-4

1. Close valves No. 123 and 124 in dust line to incinerator.
2. Close valve No. 81 air to ejector 301-4.
3. Close 8" valve No. 127 at base of dipleg and close rod out valve No. 128.
4. Open 8" valve No. 125 at top of dipleg.
5. Open aeration air valves No. 197, 198 and 199.
6. To dump cyclone dipleg:
   a. Close 8" valve No. 125 at top of dipleg.
   b. Open 8" valve No. 127 at base of dipleg to empty contents of dipleg into Dempster Dumpster box.
   c. Activate vibrator on dipleg by opening air valves No. 196 and 83. Adjusting pressure regulator valve No. 195 will vary frequency of vibrator.
   d. If necessary lower portion of dipleg can be rodded out through valve No. 128.
7. To return dust to incinerator bed:
   a. Open valve No. 123 in line from ejector to incinerator.
   b. With valve No. 124 still closed, slowly open air valve No. 81 to ejector 301-4 to establish that line is open into incinerator.
   c. Close 8" valve No. 125 at top of dipleg.
   d. Open 3" valve No. 124 at base of cyclone to admit dust to ejector 301-4.

III. EQUIPMENT OPERATING INSTRUCTIONS (Cont'd)

M. Cyclone 303-1 -- Ejector 301-4 (Cont'd)

e. This ejector air will serve as air from combustion and may necessitate a change in the setting of overbed combustion air FIC-41.

8. Cyclone dipleg will require dumping on a routine basis of once every 8 hours when the dust return system is not in service.

9. If pressure drop across cyclone increases as indicated by DPI-79 the cyclone may be plugging. Operate cyclone vibrator by opening air valves No. 82 and 196 to alleviate plugging.

N. Air Preheater 302-1

1. The fluidizing air as controlled by FIC-42 enters the incinerator via the preheater. The preheater must be operated during startups to heat the fluidizing air and thus bring the bed temperature up to a minimum of 900°F before admitting torch oil.
2. The maximum permissible outlet temperature of the preheater is 1100°F, to protect the alloy grid.
3. The preheater must not be operated until a flow of air has been established up through the incinerator bed.
4. The preheater is fired with natural gas and has the following lockouts which will shut off or prevent the flow of gas:
   a. Low fluidizing air pressure -- 3 psi -- PS-75.
   b. High natural gas pressure to burner -- 20 psi -- PS-74.
   c. Low natural gas pressure to burner -- 7 psi -- PS-73.
   d. Loss of flame -- flame sensing device.
5. Block in natural gas at battery limits when preheater is not in service to avoid popping safety valve PSV-89.
6. Pressure control valve PCV-68 should control the natural gas pressure at 25 psig as observed on pressure gage PI-68.
7. The gas pressure to the main burner is further dropped to 6" water pressure above windbox pressure by pressure control valve PCV-69 as observed on pressure gage PI-71.
8. The gas pressure to the pilot is dropped to 6" water pressure above windbox pressure by pressure control valve PCV-70 as observed on pressure gage PI-72.

III. EQUIPMENT OPERATING INSTRUCTIONS (Cont'd)

N. Air Preheater 302-1 (Cont'd)

9. The pressure indicated in 6, 7, and 8 above, must be observed with gas flowing such as when the preheater is in operation or out a vent valve.
10. To start the preheater:
    a. Check block valves No. 165, 162, and 156 open to pressure switches PS-73, PS-74, and PS-75.
    b. Pilot needle valve No. 157 is preset. Do not change setting.
    c. Check main gas control valve No. 160 closed (1-1/4" Maxon Series O control valve).
    d. Check bleed valves No. 163, 164 and bleeds on PI-71 and PI-72 closed.
    e. Establish a stable flow of at least 1100 scfm on FIC-42, fluidizing air.
    f. Open 1-1/2" gas cock No. 167 in supply line downstream of pressure control valve PCV-68.
    g. Open 1-1/2" gas cock No. 161 in main burner supply line upstream of Maxon O control valve.
    h. Open natural gas block valve at battery limits.
    i. Pilot is now ready to light: Open gas cock No. 158 in pilot supply line.
        aa. Turn "Power" switch to "On" position.
            1. The "Power On" lamp lights.
            2. The "Flame Out" lamp lights.
            3. The alarm will sound.
            4. Press "Alarm Stop" button to silence alarm.
        bb. Electric operated valve SOV-83 is still closed. Gas will not flow yet.
        cc. Depress "Start" button and hold depressed for 15 seconds or until "Flame On" lamp lights.
            1. Spark ignitor will be energized.
            2. Electric operated valve SOV-83 in pilot line will open.
        dd. When the pilot flame is sensed:
            1. "Flame On" lamp lights.
            2. "Flame Out" lamp goes out.
            3. Alarm circuit is reset.
            4. Electric valves in main burner supply line are energized so that they may be opened when ready.
        ee. If the pilot flame is not established and sensed within 15 seconds a timer will close

III. EQUIPMENT OPERATING INSTRUCTIONS (Cont'd)

N. Air Preheater 302.1 (Cont'd)

the electric operated valve SOV-83 in the pilot supply line.

ff. If the pilot does not light, or lights and goes out, the "Stop" button must be depressed to reset the timer before trying for another ignition.

gg. If the pilot does not light close gas cock No. 158 in pilot supply line and check to see that the fluidizing air pressure and gas pressures are within their specified limits for pressure switches PS-73, and PS-75, and try step "i" again.

j. To light main burner after pilot is lit:

aa. Check main gas control valve No. 160 closed (1-1/4" Maxon Series Q control valve).

bb. Open Maxon Trip Release Shut-off valve SOV-81 by turning lever clockwise to engage latch then counter-clockwise to open valve. Leave in counter-clockwise position.

cc. Electric valve SOV-82 will automatically open after opening SOV-81.

dd. Maxon Series Q Control Valve, No. 160, can now be slowly opened to its lowest setting. Observe burner flame through preheater sight-port.

ee. If main burner does not light press "Stop" button and close battery limits gas block.

11. Proceed to preheat incinerator as per incinerator operating instructions.

12. To shut down preheater fire:

a. Turn the Maxon Series Q Control Valve No. 160, to off position.

b. If it is desired to shut off the pilot:

aa. Press the "Stop" button on the Flame Control Panel and hold until alarm sounds. Press "Alarm Stop" button to silence alarm.

bb. Close the pilot cock valve No. 158, main burner cock valve No. 161, and main line cock No. 167.

cc. Close the battery limits gas block.

dd. Turn "Power" switch to "Off" position.

III. EQUIPMENT OPERATING INSTRUCTIONS (Cont'd)

O. Screw Unloading Conveyor 307-1

1. Check incinerator unloading valve No. 117 closed
2. Start screw conveyor on slow speed.
3. Open cooling water drain valve No. 118.
4. Open cooling water inlet valves No. 70 and 71.
5. Slowly open incinerator unloading valve No. 117, allowing bed material to flow into and fill conveyor.
6. Open unloading valve No. 117 wide open and control rate of discharge by varying speed of conveyor. Do not exceed 250°F outlet temperature.
7. When the desired bed level is reached in the incinerator, close unloading valve No. 117 and continue to run conveyor until bed material is cleared from conveyor.
8. Stop screw conveyor.
9. Close water inlet valves No. 70 and 71.
10. In freezing weather drain water jacket and hollow flight screw.

P. Sand Loading -- Ejector 301-3

1. Check hopper outlet valve No. 115 closed.
2. Fill hopper with sand using Lull or other lifting device. Hopper holds about 32 cubic feet or approximately two tons of sand. Incinerator requires approximately 95 cu. ft., 11,000 lbs, for a complete charge.
3. Open incinerator inlet valve No. 77 and ejector air valves No. 206 and 76 and determine if incinerator loading nozzle is free of obstruction.
4. Open hopper outlet valve No. 115 to permit sand to enter ejector 301-3. Throttle hopper outlet valve as necessary to control rate of sand addition.
5. The rate of sand addition will be determined by the ability of the bed to maintain its temperature.
6. If sand is being added to an empty incinerator, be sure aeration has been turned on to torch oil nozzles, free-board sprays, and pressure taps.
7. When shutting down sand loading, close hopper outlet valve No. 115, vessel inlet valve No. 77 and ejector air valve No. 76 in the order listed.
8. If the incinerator is in operation this loading air will act as combustion air and may necessitate an adjustment of overbed air FIC-41.

IV. FEED PREPARATION

1. When the feed mix tank 902-2 contains 900 bbls of blended incinerator feed the incinerator will be started up and kept in operation until the Feed Tank is at minimum gage. This will be a "batch" of about 650 bbls and at a feed rate of 4 gpm with 20% oil in feed, would take about 5 days to run off.
2. Sludge from the API separator slop tanks should not be pumped into the Feed Mix Tank 902-2 while the incinerator is running unless permission is obtained from the Ultraformer Operating Foreman.
3. Sludge from the API Separator Box Bottoms will be pumped each morning while the Utilities operator is skimming the API Separator Box. This stream can be pumped into the Feed Mix Tank 902-2 while the incinerator is running since it contains about 15% oil and will not significantly change the feed composition.
4. The approximate oil content of the feed components is as follows: (by volume)

API Separator Box Bottoms	10 - 20%
API Separator Slop Tank Sludge	50%
Spent Caustic	0% (no heating value)
Tank Cleaning Bottoms	± 80%

5. When a "batch" of feed is blended use the following formula to determine the approximate percent oil in the blended feed:

$$\frac{[(A)(z)]+[(b-A)(0.5)]+[(C-D)(y)]}{A+(B-A)+(C-D)+E} = \%\ \text{oil in feed}$$

A = Feed tank gage before pumping API Slop Tank Sludge. (bbls)
B = Feed tank gage after pumping API Slop Tank Sludge. (bbls)
C = Tank Bottoms gage before transferring to Feed Tank. (bbls)
D = Tank Bottoms gage after transferring to Feed Tank. (bbls)
E = Bbls of spent caustic to be blended.
z = % oil in Feed Tank before blending batch.
y = % oil in Tank Bottoms Sump before transferring to Feed Tank.

The actual percent oil in the feed is to be determined by sending a sample to the lab.

IV. FEED PREPARATION (Cont'd)

6. The denominator, (sum of quantities below the line), of the formula in paragraph 5 must not exceed the maximum gage for the Feed Tank, 900 bbls.
7. After determining the percent oil in the new feed batch, refer to Fig. 1 to determine the sludge feed rate range.
8. The optimum percent oil in the sludge feed is 25 to 30% however, the resulting percent oil will depend upon the ratios of spent caustic and oily sludges accumulated. If there is not enough spent caustic available to dilute the oily sludge, water can be pumped into the feed tank by the API Marlow pump.

V. STARTUP OF INCINERATOR

A. Preliminary Steps

1. Prepare a usable blend in the feed mix tank 902-2.
2. Turn on feed mix tank mixer 905-1. THIS MIXER MUST BE IN OPERATION WHENEVER THE INCINERATOR IS RUNNING. If the mixer fails, the incinerator operation MUST be shut down.
3. Establish that the cyclone dust line into the incinerator is open by opening block valve No. 123, checking cyclone outlet valve No. 124 closed, and opening air valve No. 81 to ejector 301-4. When flow is established, close block valve No. 123 and air valve No. 81.
4. Establish that sand loading line into incinerator is open by opening block valve No. 77, checking hopper outlet valve No. 115 closed, and opening air valve No. 76 to ejector 301-3. When flow is established, close block valve No. 77 and air valve No. 76.
5. Establish air purge flow to instrument pressure taps: above the grid FI-77, freeboard space FI-48, and downstream of cyclone FI-79, and density tap.
6. Check that cyclone dipleg is empty. Operate vibrators momentarily.
7. Check that Torch Oil tank 902-4 is full and torch oil is up to temperature 170° - 200°F.
8. Check that Sand Hopper is full of sand.
9. Open water block valves No. 169 & 170 so that emergency water valve to sludge feed line SOV-90 is lined up for operation. Check SOV-90 closed.
10. Check block valve No. 117 closed on 6" unloading line to screw conveyor.

B. Startup -- No Bed in Incinerator

After performing the "Preliminary Steps" in Part A, Section V., the following steps should be done in sequence when placing the incinerator into operation. Refer to Piping & Instrumentation drawings No. 3 & 4 for valve, instrumentation, and equipment designations. Refer to Section III of these instructions for instructions and procedures to operate specific pieces of equipment.

1. Set fluidizing air flow FIC-42 in "automatic" position at 2000 scfm.
2. Set overbed air flow FIC-41 at 0 scfm.
3. Put Scrubber into operation (See Section III, paragraph L.)
4. Line up cyclone to drop dust into dipleg. (See Section III, paragraph M)

V. STARTUP OF INCINERATOR (Cont'd)

B. Startup -- No Bed in Incinerator (Cont'd)

5. Start air compressor. (See Section III, paragraph K.)
6. Partially open and establish a steam flow through each torch oil nozzle, valves No. 141, 142, 143 & 144. Leave these steam valves cracked open. Torch oil valves No. 133, 134, 135 and 136 must be closed.
7. Establish a steam flow through both freeboard spray nozzles by opening steam valves No. 177 and 178. After determining that each spray is open, adjust steam throttle valve No. 78 to have 10 psi on pressure gage No. PI-59.
8. Establish that sludge feed line and nozzle is open by opening feed block valve No. 114 at incinerator, checking discharge block valve No. 23 at feed pump closed, and opening steam valve No. 8, located just downstream of discharge block valve on feed pump. Leave steam valve No. 8 cracked open until sludge feed is started.
9. Establish that atomizing steam nozzle, concentric with feed nozzle, is open by opening steam bypass valve No. 105 around PSV-51 control valve. Leave bypass valve cracked open until PIC-51 is placed in service.
10. Open water block valves No. 66, 67, 175, and 176 so that TRC-49, freeboard sprays, are lined up for operation. Close bypass valve No. 68 around PCV-49. Check PCV-49 control valve closed.
11. Start air preheater and bring windbox temperature to 500°F. (See Section III, paragraph N.) Hold at this temperature until all incinerator thermocouples read approximately 300°F. Control preheater outlet temperature by adjusting Maxon Series Q Control Valve No. 160.
12. Gradually increase windbox temperature TR-1 to 1100°F at a rate of approximately 100° per hour. Do not exceed 1100°F in windbox.
13. When windbox has reached 700°F, system is ready for sand loading. (See Section III, paragraph P.) Rate of sand loading will depend upon how fast sand can be heated. Throttle hopper valve No. 115 to hold bed temperature between 700°F and 1000°F in windbox. It will take approximately 95 cu. ft. of sand (11,000 lbs) to charge an empty incinerator.
14. When required amount of sand has been loaded and bed temperature is 950° to 1100°F, observe system pressure readings:

V. STARTUP OF INCINERATOR (Cont'd)

B. Startup -- No Bed in Incinerator (Cont'd)

14. (Cont'd)

	Normal	Actual
Windbox pressure PI-47	_____	_____
Grid pressure drop DPI-77	_____	_____
Bed pressure drop DPI-78	_____	_____
Cyclone pressure drop DPI-79	_____	_____
Freeboard pressure PI-48	_____	_____
Bed density	_____	_____

15. Place O_2 Analyzer 61 into operation. Check a sample with an Orsat.
16. Set freeboard sprays TRC-49 to control at 1000°F on "automatic."
17. At this point before injecting torch oil record the following temperatures:
    - TR 1 Windbox
    - TR 2 Bottom Bed
    - TR 3 Bottom Bed
    - TR 4 Bottom Bed
    - TR 5 Top Bed
    - TR 6 Middle Bed
    - TR 7 Top Bed
    - TR 8 Freeboard
    - TRC 49 Gas Outlet
    - TR 9 Freeboard

18. When bed temperature is 950° to 1100°F and bed pressure drop DPI-78 is approximately 72 inches H_2O, the torch oil can be started to bring bed up to an operating temperature of 1350°F. (See Section III, paragraph H for operation of torch oil pump.) Use the west torch oil nozzle first because this is nearest a thermocouple in the bed.
    a. Start torch oil system circulating.
    b. Open steam atomizing valve No. 141 and adjust to 15 psi.
    c. Open torch oil valve No. 133 wide open for 15 seconds, then close again.

    Watch bed termocouples for an increase in temperature. Watch oxygen analyzer O_2I-61 for a decrease in oxygen, to indicate combustion of torch oil.

V. STARTUP OF INCINERATOR (Cont'd)

B. Startup -- No Bed in Incinerator (Cont'd)

If a definite temperature increase is not noted within two minutes, increase the bed temperature another 50° and try again. Decrease the atomizing steam flow to 10 psi for the second try.

d. If a temperature rise is noted, reopen torch oil valve 2 turns open and wait 15 minutes for temperatures to stabilize. Then place the other three torch oil nozzles into service for even heating.

e. Increase bed temperature to 1350°F by opening all torch oil injection points equally. Slowly go from 1100° to 1350°F.

f. As bed temperatures increase, decrease fluidizing air FIC-42 to maintain 3 ft/sec superficial space velocity as indicated in Fig. 2. However, do not permit oxygen analyzer O_2I-61 to drop below 7%.

19. System is now ready for introduction of sludge feed:

a. Place feed atomizing steam controller PIC-51 into operation by opening block valves No. 103 and 104. Close bypass valve No. 105. Open block valve 131 to pressure switch PS-51 to activate solenoid valve SOV-51 in impulse line to PIC-51. Set PIC-51 to hold 20 psi on steam nozzle and in the "automatic" position.

b. Manually open overbed air flow control valve FCV-41 making sure that fluidizing air flow FIC-42 remains constant to maintain 3 ft/sec. Set overbed air FIC-41 for 700 SCFM and in "automatic" position. All future combustion air adjustments must be made with overbed air FIC-41. Fluidizing air flow FIC-42 must remain constant at approximately 1225 SCFM to maintain 3 ft/sec. velocity thru the bed.

c. Start sludge feed pump 901-2 and set pump speed control HIC-40 to the lowest setting. (See Section III, paragraph F.)

As sludge is introduced the bed temperature should increase and the O_2I-61 show a decrease in oxygen indicating that sludge is burning.

d. Increase sludge pump speed and simultaneously decrease first torch oil, then preheater to maintain oxygen analysis O_2I-61 at 5% or more, and bed temperature at 1350°F.

V. STARTUP OF INCINERATOR (Cont'd)

B. Startup -- No bed in Incinerator (Cont'd)

19. (Cont'd)

e. As sludge feed is increased and torch oil and gas are cut off, the bed temperature will gradually decrease. When bed temperatures stabilize, adjust feed to maintain a bed temperature of 1350°F and adjust overbed air FIC-41 to maintain oxygen O_2I-61 at 5%.

C. Startup -- Slumped Bed in Incinerator

This startup procedure assumes a normal, controlled shutdown after the previous run.

After performing the "Preliminary Steps" in paragraph A, Section V, the following steps should be done in sequence when placing the incinerator into operation. Refer to Piping & Instrumentation drawings No. 3 and 4 for valve, instrument, and equipment designations. Refer to Section III of these instructions for instructions and procedures to operate specific pieces of equipment.

1. Set fluidizing air flow FIC-42 in "manual" position and adjust for maximum flow.
2. Set overbed air flow FIC-41 closed at 0 SCFM.
3. Put scrubber into operation (See Section III, paragraph M.)
4. Line up cyclone to drop dust into dipleg. (See Section III, paragraph M.)
5. Start air compressor. (See Section III, paragraph K.)
6. If the fluidizing air flow FIC-42 is 1900 SCFM or more, the bed is sufficiently fluidized to proceed. If 1900 SCFM cannot be attained through FIC-42, consult the Operating Foreman.
7. Partially open and establish a steam flow through each torch oil nozzle, valves No. 141, 142, 143 and 144. Leave these steam valves cracked open. Torch oil valves No. 133, 134, 135 and 136 must be closed.
8. Establish a steam flow through both freeboard spray nozzles by opening steam valves No. 177 and 178. After determining that each spray is open, adjust steam throttle valve No. 78 to have 10 psi on pressure gage No. PI-59.

V. STARTUP OF INCINERATOR (Cont'd)

C. Startup -- Slumped Bed in Incinerator

9. Establish that sludge feed line and nozzle is open by opening feed block valve No. 114 at incinerator, checking discharge block valve No. 23 at feed pump closed, and opening steam valve No. 8, located just downstream of discharge block valve on feed pump. Leave steam valve No. 8 cracked open until sludge feed is started.
10. Establish that atomizing steam nozzle, concentric with feed nozzle, is open by opening steam bypass valve No. 105 around PSV-51 control valve. Leave bypass valve cracked open until PIC-51 is placed in service.
11. Open water block valves No. 66, 67, 175 and 176 so that TRC-49, freeboard sprays, are lined up for operation. Close bypass valve No. 68 around PCV-49. Check PCV-49 control valve closed.
12. Start air preheater and increase bed temperature 100° per hour until bed reaches 950°F to 1100°F. Do not exceed 1100°F in windbox. Control preheater outlet temperature by adjusting Maxon Series Q control valve No. 160. Record temperatures as follows:

TR 1	Windbox	_____
TR 2	Bottom Bed	_____
TR 3	Bottom Bed	_____
TR 4	Bottom Bed	_____
TR 5	Top Bed	_____
TR 6	Middle Bed	_____
TR 7	Top Bed	_____
TR 8	Freeboard	_____
TR 9	Gas Outlet	_____
TRC 49	Freeboard	_____

13. As bed temperature increases, decrease fluidizing air flow FIC-42 to maintain 3 ft/sec superficial space velocity in bed area. See Fig. 2.
14. When bed temperature is 950° to 1100°F, observe system pressure readings:

	Normal	Actual
Windbox pressure PI-47	_____	_____
Grid pressure drop DPI-77	_____	_____
Bed pressure drop DPI-78	_____	_____
Cyclone pressure drop DPI-79	_____	_____
Freeboard pressure PI-48	_____	_____
Bed density	_____	_____

V. STARTUP OF INCINERATOR (Cont'd)

C. Startup -- Slumped Bed in Incinerator (Cont'd)

15. If bed pressure drop DPI-78 and density indicate less than 6 ft. of fluidized bed, add sand to bring bed up to 6 ft. in fluidized state. (See Section III, paragraph P.)
16. Place O_2 Analyzer 61 into operation. Check a sample with an Orsat.
17. Set freeboard sprays TRC-49 to control at 1000°F on "automatic".
18. At this point, before injecting torch oil, record the following temperatures:

TR	1	Windbox	_____
TR	2	Bottom Bed	_____
TR	3	Bottom Bed	_____
TR	4	Bottom Bed	_____
TR	5	Top Bed	_____
TR	6	Middle Bed	_____
TR	7	Top Bed	_____
TR	8	Freeboard	_____
TR	9	Gas Outlet	_____
TRC	49	Freeboard	_____

19. When bed temperature is 950°F to 1100°F and bed pressure drop DPI-78 is approximately 72 inches H_2O, the torch oil can be started to bring bed up to an operating temperature of 1350°F. (See Section III, paragraph H for operation of torch oil pump.) Use the west torch oil nozzle first because this is nearest a thermocouple in the bed.
    a. Start torch oil system circulating.
    b. Open steam atomizing valve No. 141 and adjust to 15 psi.
    c. Open torch oil valve No. 133 wide open for 15 seconds, then close again.

    Watch bed thermocouples for an increase in temperature. Watch oxygen analyzer O_2I-61 for a decrease in oxygen, to indicate combustion of torch oil.

    If a definite temperature increase is not noted within two minutes, increase the bed temperature another 50° and try again. Decrease the atomizing steam flow 10 psi for the second try.

V. STARTUP OF INCINERATOR (Cont'd)

C. Startup -- Slumped Bed in Incinerator (Cont'd)

d. If a temperature rise is noted, reopen torch oil valve 2 turns open and wait 15 minutes for temperatures to stabilize. Then place the other three torch oil nozzles into service for even heating.
e. Increase bed temperature to 1350°F by opening all torch oil injection points equally. Slowly go from 1100° to 1350°F.
f. As bed temperatures increase, decrease fluidizing air FIC-42 to maintain 3 ft/sec superficial space velocity as indicated in Fig. 2. However, do not permit oxygen analyzer O_2I-61 to drop below 7%.

20. System is now ready for introduction of sludge feed:
a. Place feed atomizing steam controller PIC-51 into operation by opening block valves No. 103 and 104. Close bypass valve No. 105. Open block valve 131 to pressure switch PS-51 to activate solenoid valve SOV-51 in impulse line to PIC-51. Set PIC-51 to hold 20 psi on steam nozzle and in the "automatic" position.
b. Manually open overbed air flow control valve FCV-41 making sure that fluidizing air flow FIC-42 remains constant to maintain 3 ft/sec. Set overbed air FIC-41 for 700 SCFM and in "automatic" position.

All future combustion air adjustments must be made with overbed air FIC-41. Fluidizing air flow FIC-42 must remain constant at approximately 1225 SCFM to maintain 3 ft/sec. velocity thru the bed.

c. Start sludge feed pump 001-2 and set pump speed control HIC-40 to the lowest setting. (See Section III, paragraph F.)

As sludge is introduced, the bed temperature should increase and the O_2I-61 show a decrease in oxygen indicating that sludge is burning.

d. Increase sludge pump speed and simultaneously decrease first torch oil, then preheater to maintain oxygen analysis O_2I-61 at 5% or more, and bed temperature at 1350°F.
e. As sludge feed is increased and torch oil and gas is cut off, the bed temperature will gradually decrease. When bed temperatures stabilize, adjust

V. STARTUP OF INCINERATOR (Cont'd)

C. Startup -- Slumped Bed in Incinerator (Cont'd)

feed to maintain a bed temperature of 1350°F and adjust overbed air FIC-41 to maintain oxygen O_2I-61 at 5%.

VI. SHUTDOWN OF INCINERATOR

A. Normal Shutdown

1. Shutdown sludge feed pump 901-2 by shutting off motor.
2. Close sludge block valve No. 23 at pump discharge.
3. Increase overbed air flow FIC-41 to maximum.
4. Crack open steam block valve No. 8 at pump discharge ,and slowly blow sludge feed line into incinerator.
5. After sludge line is steam purged, decrease overbed air FIC-41 to one-half line. Do not drop below 5 percent oxygen.
6. Close steam purge valve No. 8 and close sludge feed block valve No. 114 at feed nozzle.
7. Reduce atomizing steam to feed nozzle, PIC-51, to a minimum for purge flow only.
8. Gradually increase fluidizing air flow FIC-42 as bed cools down to maintain a space velocity of 3 ft/sec in bed. Refer to Fig. 2.
9. If bed level has increased to above 72 inches during this run, unload to proper bed level. (See Section III, paragraph O.)
10. Dump cyclone leg. (See Section III, paragraph M.)
11. Discontinue cyclone dust to incinerator if in operation. (See Section III, paragraph M.)
12. Continue fluidizing air flow until bed temperature cools to 700°F, then shut down air compressor. (See Section III, paragraph K.)
13. Shut down scrubber. (See Section III, paragraph L.)
14. Shut off steam purges to torch oil nozzles; feed nozzle, spray nozzles when scrubber is shut down. Leave air purges to level instruments in operation.

Sensible heat will dissipate in 2 or 3 days.

B. Emergency Shutdown -- Air Compressor or Power Failure

Failure of the air compressor or a total electrical failure will cause an emergency shutdown of the incinerator.

The feed pump and torch oil pump will shut off automatically.

The air preheater gas solenoid valves will close and the preheater will shut down automatically.

The feed atomizing steam will continue to flow and tend to purge the incinerator vessel.

VI. SHUTDOWN OF INCINERATOR (cont'd)

B. Emergency Shutdown -- Air Compressor or Power Failure (cont'd)

Perform the following steps when time permits:

1. Block in feed line at the nozzle. Do not blow out with steam.
2. Sand loading ejector air valve No. 76 and cyclone dust return ejector air valve No. 81 must be closed if either system were in service.
3. Close all torch oil nozzle valves if any were in service.
4. Block in natural gas battery limits block valve.
5. Leave all nozzle purges, both air and steam, in operation until bed has cooled to 400°F.

   Bed will defluidize and slump at operating temperature and any chemicals present may cause the bed to fuse. This "fused" bed may have to be removed manually before incinerator can be restarted.
6. Shutdown scrubber when flue gas temperature has dropped to 600°F. (See Section III, paragraph L.)

C. Emergency Shutdown -- Steam Failure

Steam failure will cause loss of feed atomizing steam and will necessitate an immediate near normal shutdown of the incinerator.

1. Reduce feed to a minimum with HIC-40.

Perform the following steps when time permits.

2. Stop sludge feed pump.
3. Close sludge feed block valve No. 114 at feed nozzle.
4. Close steam block valve No. 104 at atomizing steam control valve PIC-51.
5. Close steam throttle valves No. 141, 142, 143, and 144 at torch oil nozzles.
6. Close steam block valves No. 177 and 178 at freeboard water spray.
7. Decrease overbed air FIC-41 to one-half line. Do not drop below 5 percent oxygen.
8. If bed level was increased to above 72 inches during this run, unload to proper bed level. (See Section III, paragraph O.)
9. Dump cyclone leg. (See Section III, paragraph M.)
10. Discontinue cyclone dust to incinerator if in operation. (See Section III, paragraph M.)

VI. SHUTDOWN OF INCINERATOR (cont'd)

C. Emergency Shutdown -- Steam Failure (cont'd)

11. Continue fluidizing air flow until bed temperature cools to 700°F, then shut down air compressor. (See Section III, paragraph K.)
12. Shut down scrubber. (See Section III, paragraph L.)

VII. NORMAL OPERATION VII-1

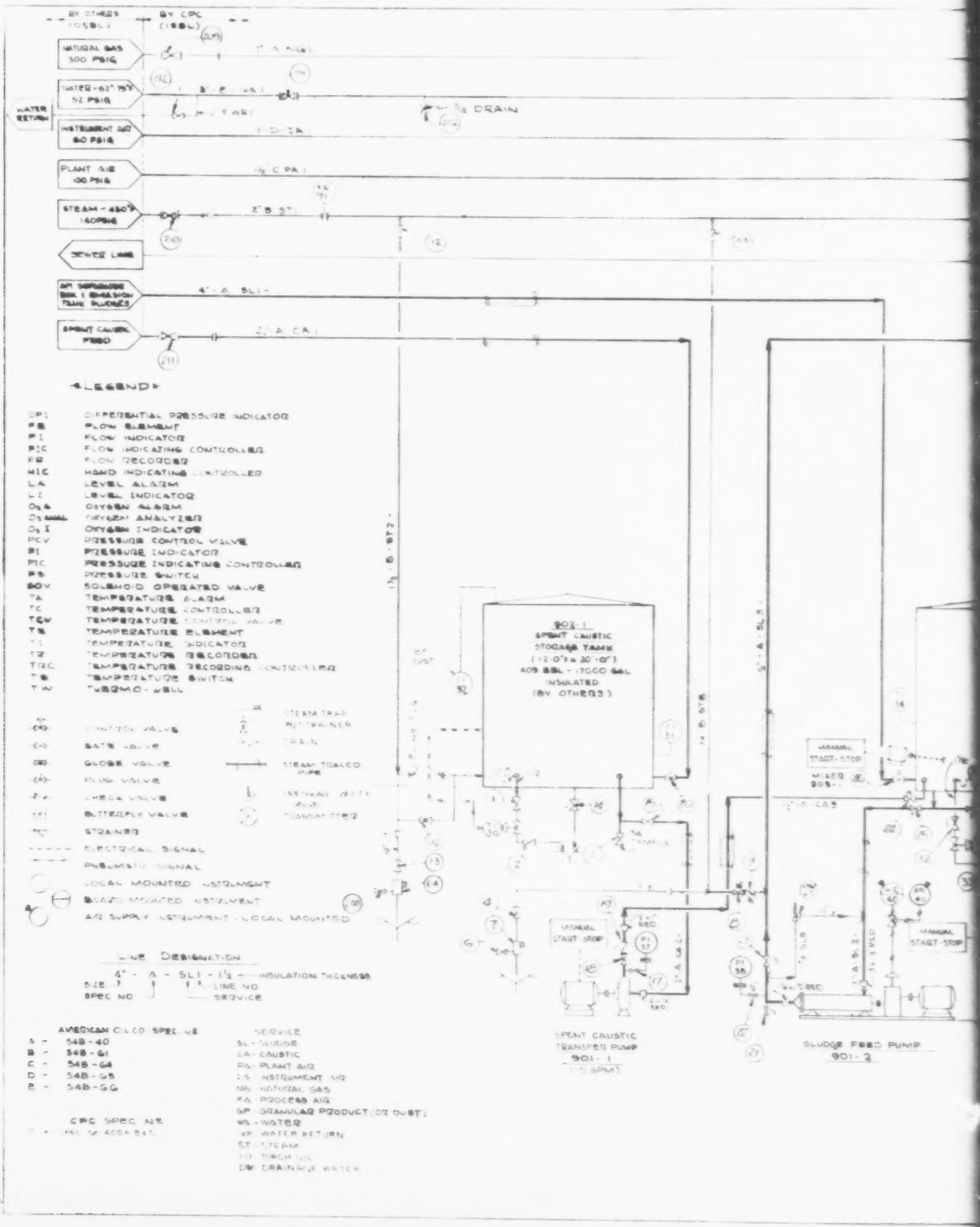
After startup of the system as described in Section V, the operation should continue until the feed tank has been pulled down to its minimum gage.

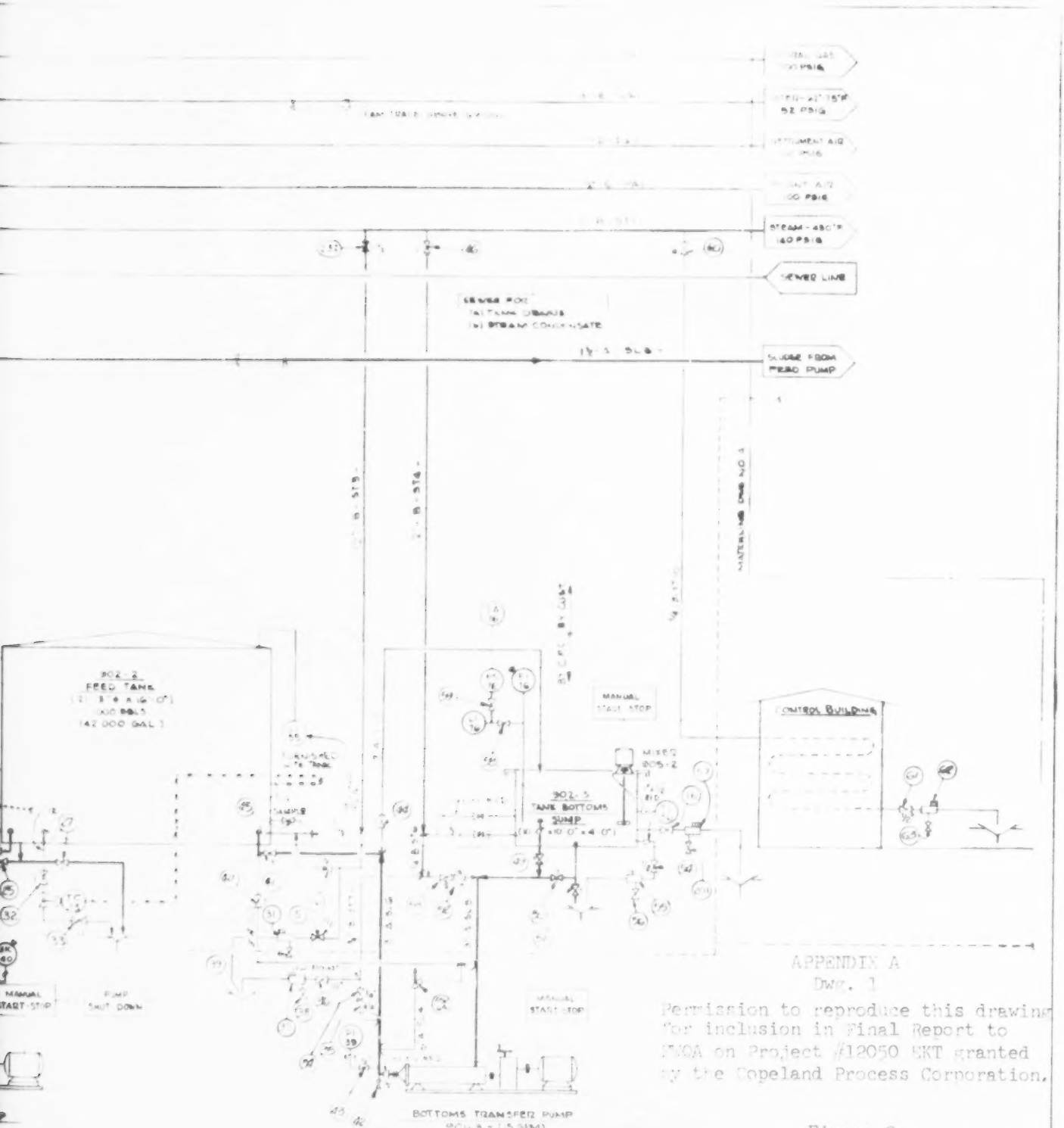
1. Operate at maximum feed rate consistent with the fusion properties of the bed and the air available for combustion.
2. If the bed level builds more than 12" after some extended period of operation, operate the screw conveyor to remove sufficient material to establish the proper level. (See Section III, paragraph O.) Whenever bed material is removed, send a 1 quart sample to the lab for analysis.
3. If the cyclone fines are not being returned to the incinerator, the cyclone dipleg must be emptied once per 8-hour shift. (See Section III, paragraph M.)
4. If the bed level falls more than 12", add sand to the incinerator to restore the operating level of the bed. (See Section III, paragraph P.)
5. If sampling indicates poor particle size distribution of bed material, unload some bed material and add some fresh sand to restore particle size distribution and improve fluidization performance. A desirable distribution of particle size is as follows:

U.S. Sieve Size		Wt. %
	+ 9	2
- 9	+20	10-35
-20	+35	15-35
-35	+48	15-35
-48	+60	8-20
-60		2-10

VII. NORMAL OPERATION (cont'd)

6. If sampling indicates a need for fines, the cyclone dust return system should be placed in operation. (See Section III, paragraph M.)
7. As the bed material becomes coated with chemicals and the small particles agglomerate, the fluidization properties of the bed will deteriorate and show up as a loss or fluctuation in differential bed pressure DPI-78. This may be remedied by returning cyclone fines to the incinerator or replacing a portion of the bed with new sand.
8. If a chemical bed is operated at too high a bed temperature, melting occurs, quickly followed by de-fluidization and fusing of the bed.
9. If the freeboard temperature, with a chemical bed, is operated at too high a temperature, the dust particles will be sticky and cause plugging of the cyclone and ducts.
10. If the mixer on the Feed Mix Tank fails, the incinerator must be shut down.


BY OTHERS (105BL)
BY CPC (198BL)
NATURAL GAS 300 PSIG
WATER - 62° 75% 52 PSIG
WATER RETURN
INSTRUMENT AIR 80 PSIG
PLANT AIR 100 PSIG
STEAM - 480° 140PSIG
SEWER LINE
API SEPARATOR SLUDGE / EMULSION TANK SLUDGES
SPENT CAUSTIC FEED
DRAIN
LEGEND
DPI DIFFERENTIAL PRESSURE INDICATOR
FE FLOW ELEMENT
FI FLOW INDICATOR
FIC FLOW INDICATING CONTROLLER
FR FLOW RECORDER
HIC HAND INDICATING CONTROLLER
LA LEVEL ALARM
LI LEVEL INDICATOR
O2 A OXYGEN ALARM
O2 ANAL OXYGEN ANALYZER
O2 I OXYGEN INDICATOR
PCV PRESSURE CONTROL VALVE
PI PRESSURE INDICATOR
PIC PRESSURE INDICATING CONTROLLER
PS PRESSURE SWITCH
SOV SOLENOID OPERATED VALVE
TA TEMPERATURE ALARM
TC TEMPERATURE CONTROLLER
TCV TEMPERATURE CONTROL VALVE
TE TEMPERATURE ELEMENT
TI TEMPERATURE INDICATOR
TR TEMPERATURE RECORDER
TRC TEMPERATURE RECORDING CONTROLLER
TS TEMPERATURE SWITCH
TW THERMO-WELL
CONTROL VALVE
GATE VALVE
GLOBE VALVE
PLUG VALVE
CHECK VALVE
BUTTERFLY VALVE
STRAINER
ELECTRICAL SIGNAL
PNEUMATIC SIGNAL
LOCAL MOUNTED INSTRUMENT
BOARD MOUNTED INSTRUMENT
AIR SUPPLY INSTRUMENT - LOCAL MOUNTED
STEAM TRAP W/STRAINER
DRAIN
STEAM TRACED PIPE
TRANSMITTER
LINE DESIGNATION
4" - A - SL1 - 1½ — INSULATION THICKNESS
SIZE
SPEC NO
LINE NO
SERVICE
AMERICAN OILCO SPEC NO
A - 54B-40
B - 54B-61
C - 54B-64
D - 54B-65
E - 54B-66
CPC SPEC NO
SERVICE
SL - SLUDGE
CA - CAUSTIC
PA - PLANT AIR
IA - INSTRUMENT AIR
NG - NATURAL GAS
PA - PROCESS AIR
SP - GRANULAR PRODUCT (OR DUST)
WS - WATER
WR - WATER RETURN
ST - STEAM
FO - FUEL OIL
DW - DRAINAGE WATER
901-1 SPENT CAUSTIC STORAGE TANK (12'-0" ø x 20'-0") 409 BBL - 17000 GAL INSULATED (BY OTHERS)
SPENT CAUSTIC TRANSFER PUMP 901-1
SLUDGE FEED PUMP 901-2
MANUAL START-STOP
MIXER 901-1



APPENDIX A
Dwg. 1

Permission to reproduce this drawing for inclusion in Final Report to EPA on Project #12050 EKT granted by the Copeland Process Corporation.

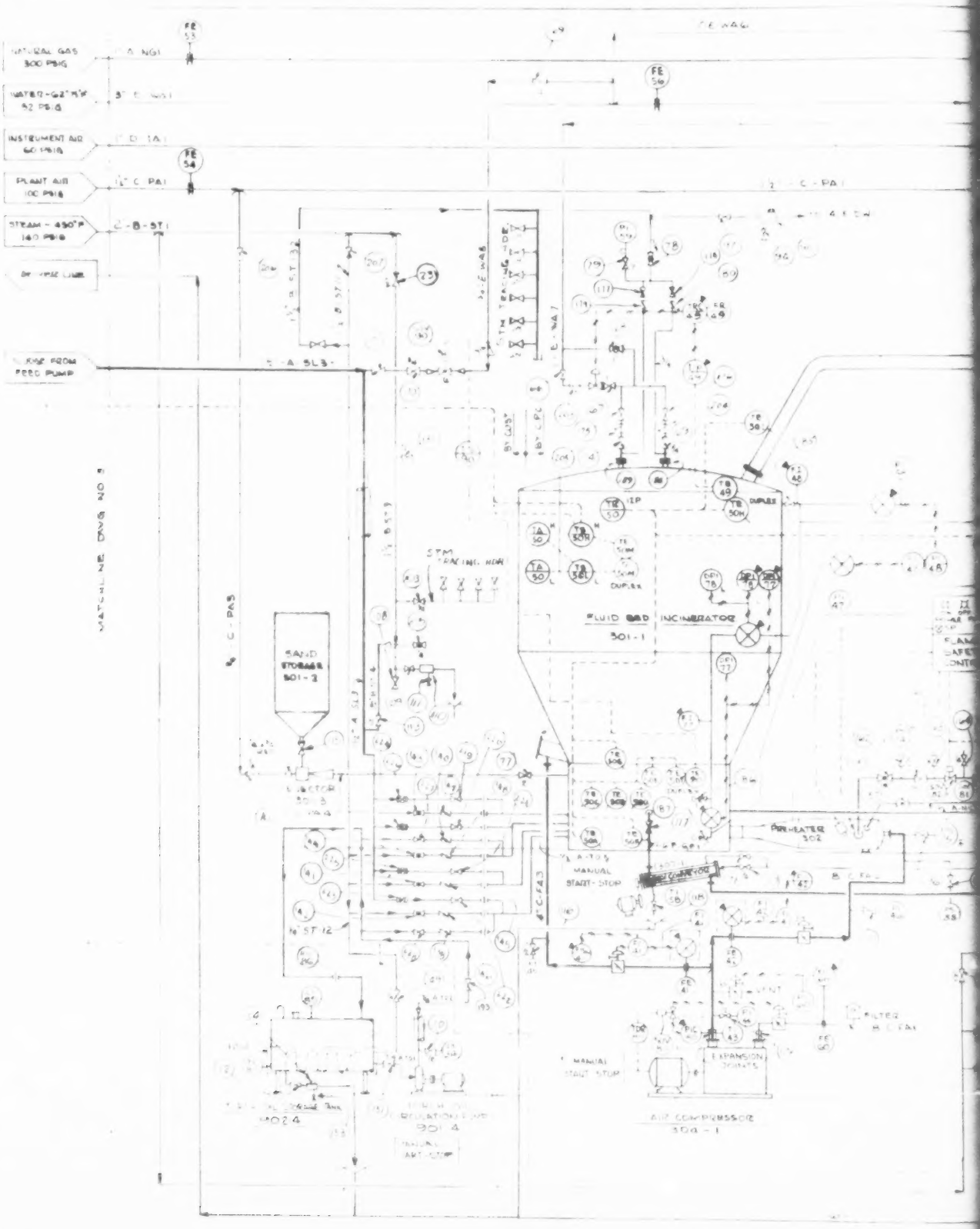
Figure 9

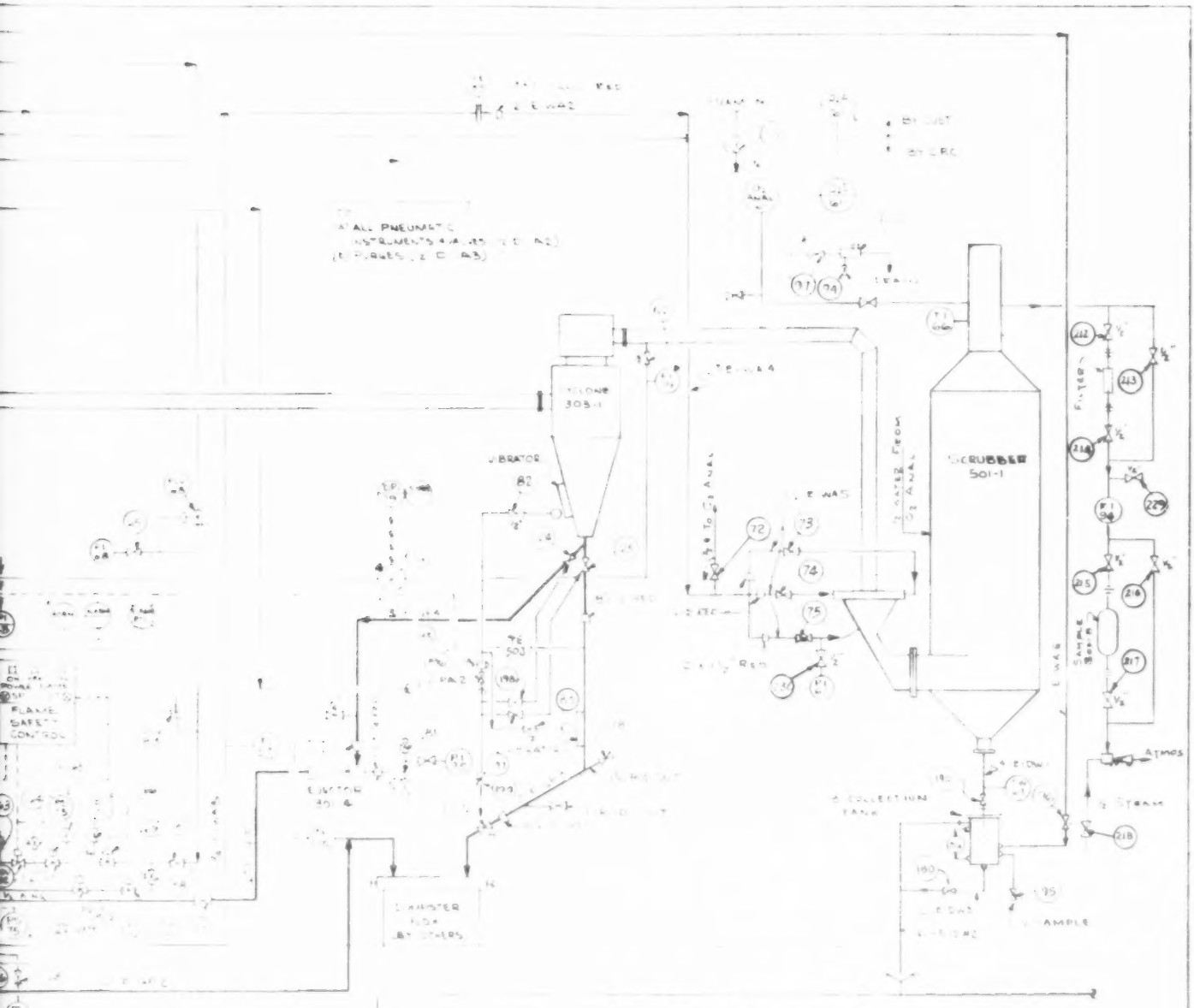
COPELAND PROCESS CORPORATION
120 OAKBROOK MALL
OAKBROOK, ILLINOIS

CPC

AMERICAN OIL COMPANY
MANDAN, N. D.

PIPING & INSTRUMENTATION DIAGRAM
FOR
OILY SLUDGE & SPENT CAUSTIC INCINERATOR


NATURAL GAS 300 PSIG
WATER 62°F 95# 92 PSIG
INSTRUMENT AIR 60 PSIG
PLANT AIR 100 PSIG
STEAM - 450°F 140 PSIG
BY-PASS LINE
FROM FEED PUMP
FE 53
FE 54
FE 56
FE 49
TE 49
TE 50
TE 50H
TE 53
TA 50
TA 50H
TA 50L
DUPLEX
IP
STM TRACING HDR
FLUID BED INCINERATOR
301-1
SAND STORAGE 301-2
PREHEATER 302
MANUAL START-STOP
EXPANSION JOINTS
AIR COMPRESSOR 304-1
REGULATOR 301-4
304-4
FILTER
VENT



APPENDIX A
Dwg. 2

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Copeland Process Corporation.

Figure 10

COPELAND PROCESS CORPORATION
120 OAKBROOK MALL
OAKBROOK, ILLINOIS

CPC

AMERICAN OIL COMPANY
MANDAN, N.D.

PIPING & INSTRUMENTATION DIAGRAM
FOR
OILY SLUDGE & SPENT CAUSTIC INCINERATOR

DRAWN JN16
SHEET NO. 4

B

APPENDIX B

CHRONOLOGICAL LOG OF INCINERATOR OPERATIONS JUNE 20, 1969 - JANUARY 30, 1970

TRAINING

Prior to startup operations, ten operators and three first line supervisors received 10 hours of classroom and field training (conducted by American Oil Co. personnel). Following this training, American Oil Co. operators assisted Copeland with startup attempts.

OPERATIONS

June 20, 1969

Checked operation of equipment including air compressor and air preheater. Fired air preheater June 18. Held wind box temperature at 750°F. thru June 20 to cure incinerator refractory lining. Shut unit down to make minor refractory repairs.

June 23, 1969

Startup operations began 9 a.m. 30 bags of sand loaded to provide a bed delta pressure of 62 in water. Wind box temperature raised to 840°F. by firing the air preheater. Bed temperature reached only 540°F. As a result of excessive heat loss from the wind box, it was not possible to get bed temperature high enough to introduce torch oil. Unit was shut down at 8 p.m. June 23. Additional interior insulation was applied in the wind box and a spring was changed in the fuel gas regulator for better preheater firing control.

June 28, 1969

Startup operations began at 11:45 a.m. June 26. By 8 a.m. June 27, sand was reloaded and the wind box temperature was up to 1050°F. The outside wind box skin temperature reached 640°F. as indicated by contact pyrometer readings. With wind box temperature 1100°F, the bed temp reached 825°F. The wind box temperature was raised

to 1200°F., and when the bed temperature reached 880°F. torch oil was introduced. Sludge feed was introduced about 6 p.m. Saturday June 28. The feed pump speed changer was found to be stuck in the slow position. Because of inability to control feed, the unit was shut down. The combustion air was shut down as soon as the bed temp fell to 1100°F., thus conserving heat for the next startup.

June 30, 1969

The air compressor and preheater were started up at 9:25 a.m. June 29. By 10 a.m. the bed temperature was high enough to permit the introduction of torch oil. Sludge feed was admitted at 1 p.m. and the torch oil was shut off at 3 p.m. When the sludge was admitted, the free board temperature (above the bed) reached 1500°F. with the water quench valve about 1/2 open, however it was necessary to continue firing the air preheater maintaining a wind box temperature of 930°F. to prevent the bed temperature from falling below combustion temperature. Atomizing steam to the sludge burning nozzle was shut off and the bed temperature rose 100 degrees from 1100 to 1200°F. At 6 a.m. on June 30 the dust return line between the cyclone and the incinerator broke necessitating a shutdown to repair the line.

General Comments

While sludge feed is in the unit, it has thus far been impossible to maintain the bed temperature with the air preheater shut down. Since the sludge is introduced above the bed it is believed that rapid vaporization and combustion occurs above the bed, thus permitting the bed temperature to fall while the free board temperature rises. It is planned to lower the sludge feed location and to increase the bed depth.

July 2, 1969

Dust return line between the cyclone and the incinerator repaired. Air preheater fired 8:20 a.m. Torch oil to unit at 2:30 p.m. Sludge feed charged shortly thereafter. Electrical failure occurred 9:30 p.m. when unit lights were turned on. Unit was shut down.

July 9, 1969

Electrical repair work completed July 7, 1969. Combustion air blower started 9:30 a.m. July 8. Air preheater fired 10 a.m. Skin temperature on wind box shell reached 850°F. with sand bed at 860°F. Torch oil was injected and sludge introduced between 10:30 and 11:00 p.m. Air preheater off at 11:30 p.m. Bed temperature could not be maintained with sludge feed. Torch oil used intermittently. 60 Gal. torch oil charged between midnight and 8:00 a.m. July 9.

Shell of wind box discovered cracked in three places. Thermal stress from high shell temperature from firing air heater apparently caused cracks. Incinerator was shut down for grid support design change.

August 18, 1969

Grid support design change work completed. Drying out refractory lining in preparation for startup.

August 21, 1969

Copeland fluidized bed incinerator has been in operation three days this week after refractory dryout the first two days of the week. Several operating problems remain:

1) Cannot maintain sufficiently high bed temperature to sustain combustion without torch oil.
2) Gas preheater kicks off unexplainedly but has not prevented startup.
3) Shutdown circuit interlock has to be bypassed until burned off thermocouples are replaced and rerouted.

The grid changes appear to have solved the shell overheating. A windbox temperature of 1150°F. results in a maximum shell temperature of 270°F. Minor problems continue in piping and valve erosion of the sand eductor and cyclone return systems.

Operations are planned Monday through Friday next week.

August 26, 1969

Copeland fluidized bed incinerator has not been operational since Friday a.m. 8-22-69 when it was shut down due to a hole which

August 26, 1969 (Cont'd)

eroded thru the body of a 3" valve in the dust return line.

While the valve was being repaired, the following was done:

1) The feed tank mixer was shut off to permit any free water to settle out. About 250 Bbls. of water was drained to the sewer. It is hoped that this enriched sludge feed will sustain bed temperature without torch oil.
2) The conduit carrying the high freeboard temperature shutdown circuit was rerouted and made operational.
3) The low gas pressure shutdown switch on the preheater was rewired.

It is planned to start up late today, Tuesday 8-26-69 and operate thru Friday 8-29-69 on oil sludge. No caustic has been charged to the incinerator as yet.

August 29, 1969

Copeland fluidized bed incinerator has operated satisfactorily without supplemental fuel since Tuesday 8-26-69. The sludge feed contains 50% oil and no spent caustic.

Plans are to shut down today, 8-29-69, and install temporary piping to inject spent caustic into the incinerator just above the bed. The unit will be started up the middle of next week and tried with some caustic injection.

September 4, 1969

Temporary piping and facilities to inject spent caustic into the original feed nozzle just above the bed have been installed. Incinerator was started up on Tuesday p.m. September 2, 1969 with a sludge feed containing 45% oil. First attempts to inject spent caustic caused drastic bed temperature changes, but subsequent close control of caustic flow has resulted in satisfactory operation with a bed temperature of 1250°F. to 1300°F. The ratio of sludge feed to spent caustic is about four to one. Some trouble is being experienced with plugging of the sludge feed pump suction screen and also the spent caustic injection nozzle.

September 4, 1969 (Cont'd)

Continued operation at the present feed rates is planned to observe the effects of the spent caustic injection.

September 8, 1969

Copeland fluidized bed incinerator was shut down late on 9-15-69 due to plugging of the caustic injection nozzle and line. The north and west feed nozzles were also plugged. In addition, the vessel nozzle through which the caustic injection nozzle is mounted was found plugged at the inside and with a hard deposit, presumably from the caustic.

Cleaning of plugged equipment and revision of the caustic injection nozzle to extend about 2-1/2 inches beyond the edge of the fire brick has been completed. Plans are to start up the incinerator today and re-establish operation with sludge containing 45% oil, with maximum injection of spent caustic.

September 12, 1969

The Copeland incinerator has operated without major difficulty since startup late Tuesday, 9-9-69, using sludge feed containing 42% oil. Sludge feed rate has been limited by air availability. Spent caustic injection just above the bed has been maximized as limited by bed temperature. Since startup this week, sludge feed rate has averaged about 2.8 barrels per hour, with 0.56 barrels per hour average caustic injection. Continued operation at maximum sludge and caustic rates is planned for the next several days.

Overbed water spray nozzles are being replaced today with wider angle nozzles in an attempt to reduce the effect of overbed spray water on bed temperatures.

September 16, 1969

Copeland incinerator operations continued without major problems. On 9-15-69 an estimated 550 Bbl. of sludge was transferred to feed tank while charging sludge and spent caustic to incinerator. Operations remained smooth during the transfer. Resulting feed mix contains 36% oil.

September 16, 1969 (Cont'd)

Installation of wider angle spray nozzles in freeboard (overbed) water sprays delayed until 9-15. Wider angle sprays markedly improved operations. Bed temperature rose from 1200 to 1380°F. permitting higher caustic feed rate.

Currently operating at a maximum spent caustic rate of about 3 BPH as limited by caustic pump capacity. Bed temperature has been reduced to 1200°F. by reducing sludge feed rate.

Bed is rapidly becoming chemical. Bed level increased about 40% in 24 hours from 9 a.m. 9-15 to 9 a.m. 9-16. Sample of bed taken this morning fused at 1750°F. in laboratory. This is a distinct change in bed characteristic from prior sample on caustic operation.

September 19, 1969

Copeland incinerator was shut down at 11 p.m. 9-18. Prior to shutdown sludge/spent caustic ratio was increased from 1/1 to about 2/1. Bed particle size distributary showed marked increase in coarser particles. Fraction retained on 20 mesh sieve increased from 1% on 9/11 (after 2 days operation at low caustic rates) to 60% on 9-17. Fraction passing 40 mesh decreased from 28% to 1% for above dates. Sample of bed taken 9-18 was 74% water soluble. Last night lower bed temperatures dropped rapidly to about 300°F. Lower bed pressure drop was only other anomaly observed initially. Bed pressure drop continued to decrease. Two hours later the temperature at the top of the bed started to drop. Orderly shutdown followed.

Plans are to dump bed and inspect interior of incinerator early next week. Startup is planned for mid-week.

September 23, 1969

On opening the Copeland incinerator, a large amount of agglomerated bed material was found. The agglomerates ranged in size from one-half inch to one foot in diameter. There was an extremely heavy buildup of material on the upper section of the incinerator walls. These stalactites and the bed are being removed. Startup planned for later in the week.

September 26, 1969

Copeland incinerator startup delayed by finding of extensive damage to the grid plate distributor pipe tees. Apparent cause of damage was the falling of the 4-section reactor manway refractory plug into the bed during the last run, with subsequent bumping of the plug sections on top of the pipe tees in the fluidized bed. Startup now scheduled for early next week. Facilities being revised to permit caustic injection at two points above the bed, instead of one, and to permit steam attrition of the bed near the top of the bed.

September 30, 1969

Copeland incinerator repairs expected to be completed late today. These include a more positive means of holding the manway refractory plug in place and straightening of the aeration tees damaged when this plug became dislodged during the last run.

Startup scheduled for tomorrow.

October 7, 1969

Startup of the Copeland incinerator began on October 1. Operations appeared satisfactory although first attempts to use air in the overhead sprays to improve atomization of the spray water was not successful. The spray water was partially backed out by the air.

The unit was shut down on Oct. 4 due to plugging of both overhead sprays, malfunctioning of the oxygen analyzer, and the apparent buildup of solids in the cyclone cone section.

Repairs to the incinerator are expected to be completed late today or early tomorrow, with the unit startup to follow shortly thereafter.

October 14, 1969

Copeland incinerator operations appear reasonably satisfactory, but having difficulty with plugging of the bed dump line, the fines return line, and sundry small problems.

Copeland incinerator shut down on 10-13-69. Apparently localized overheating in or near the bed unloading nozzle caused some fusion

October 14, 1969 (Cont'd)

and prevented bed withdrawal. The fines return line was plugged near the bottom of the dust cyclone by compaction of damp fines. The operation of the overhead sprays appears to be very sensitive with control of after burning and relationship between bed and freeboard temperatures difficult to control.

Startup planned for tonight.

October 20, 1969

Incinerator operations appear satisfactory with the sludge to caustic ratio about 3 to 1. The sludge rate is being kept at about 3 Bbls. per hour to keep the superficial space velocity reasonably low. We are having trouble with the bed unloading screw conveyor. There may be a leak from the cooling jacket causing the conveyor to clog with sticky discard bed.

October 22, 1969

Incinerator operations continue to be satisfactory except having difficulty keeping the sludge to caustic ratio consistently as low as 3 to 1. It is hoped that different overhead water spray nozzles now on order will improve this condition. The bed unloading screw conveyor has been taken out of service to test for leaks in the cooling water jacket and to make repairs to the upper bearing retainer which was broken when the conveyor plugged. Bed level is being controlled by dumping directly to barrels without precooling.

October 24, 1969

Copeland incinerator was shut down for inspection on Wednesday, October 22, preparatory to the dedication scheduled for Friday, October 31. The upper walls of the reactor were essentially clean. There was a localized heavy buildup in the area of the south caustic injection point, although the nozzle was still open. There was a small line of agglomerated bed material under each of the sludge feed nozzles. The bed particle size was increasing but still under control at the time of shutdown. It appears it will be necessary to attrit the bed on a routine basis if larger ratios of caustic to sludge are to be handled.

Startup planned for Monday night and Tuesday.

October 28, 1969

Copeland incinerator started up today. Operations appear satisfactory. Messrs. Wright and Horn from Shell at Houston were here today to observe the startup operations.

November 10, 1969

The incinerator operations continue to be relatively smooth with occasional upsets caused by such things as plugged sludge feed nozzles. The formal dedication was held today with a goodly representation of state and local officials and other interested personnel present.

November 14, 1969

The incinerator was shut down last week to accumulate sludge feed for performance testing. This feed was composed of old additives sent to us for disposal and skimmings from an old waste sludge pit. The unit was started up Wednesday, but operations have been erratic because of miscellaneous problems. It appears that the injection of air at the sludge feed nozzles for feed atomization greatly improves efficiency by reducing afterburning, thereby backing out spray water and improving the caustic to sludge ratio.

A performance test on the equipment is now planned for next week.

November 21, 1969

The incinerator was shut down on Sunday due to high cyclone pressure drop caused by buildup in the reactor outlet and in the duct to the cyclone. This plugging occurred while incinerating about two barrels per hour of caustic. The cyclone pressure drop has remained in control since starting up on Tuesday, but the caustic rate has been curtailed. A performance test was made on Wednesday and the results are still being evaluated by Technical Service.

Bed particle size continued to increase during the week in spite of adding attrition steam near the top of the bed. The bed started to slump last night so both sludge and caustic feed were cut out. We are now attempting to reduce particle size while preheating the bed and injecting steam near the bottom.

November 26, 1969

The incinerator bed was successfully attrited to a normal particle size last Friday by injecting steam through the sludge feed nozzles for five hours.

High pressure drop across the cyclone was again experienced with a buildup of deposits at the reactor outlet and in the duct to the cyclone when incinerating one and one half barrels of caustic per hour. The unit was shut down Sunday and the deposits cleaned out on Monday. We are now injecting most of the caustic at a sludge feed nozzle which is low in the bed to better desiccate the caustic and hopefully prevent the stickiness which has apparently been causing the fouling near the cyclone inlet.

Since Friday is a refinery holiday, there will be no wire then.

December 5, 1969

The incinerator was shut down on Tuesday due to a low sludge feed inventory. We are now accumulating a sludge feed similar to summer production for use in further performance testing. This sludge is being hauled to the incinerator from one of the API separator bottoms pits. The last run on the incinerator was apparently the most successful to date with no cyclone plugging observed. Injecting attrition steam through one bottom feed nozzle while the incinerator is on stream, caused the bed particle size to remain in the operational range. This will help promote more continuous operation although it does cost caustic burning capacity. Plan to start up next week after accumulating an adequate supply of feed.

December 15, 1969

We were unable to synthesize a design summer sludge feed for the incinerator because of the low oil content of the accessible sludge pit. While combustion of the resulting feed containing 17% oil, 27% sediment and 56% water was demonstrated, the last run was terminated after one day because of recurring problems with plugging of the feed pump suction. Sediment content of the feed was diluted and the oil content was increased by the addition of separator emulsion and the unit started up again on Friday. Operations have been erratic since then due to the low capacity of the feed pump.

December 22, 1969

The incinerator has been down since Monday while waiting for a stator assembly for the sludge feed pump. Rubber liner of the original stator was somewhat eroded and this restricted pump capacity, so that operations could not be sustained without torch oil. The replacement stator is being installed today with startup scheduled for later in the day.

December 26, 1969

Incinerator startup operation on 12-19-69 was unsuccessful due to uneven bed temperatures, restricted flow of fluidizing air with high pressure differential on the distributor place, and glowing-red hot spots on two locations on the west side of the shell. Unit down all week for repairs. Large clinker found in the bed across the west side, covering the west feed nozzle. Startup planned for early next week.

January 5, 1970

We have been unsuccessful in efforts to start up the incinerator using retained chemical bed. About one-half the bed was filled through the manway to speed up loading. When the blower was started the next day, the bed had caked to the point where only a small amount of fluidizing air could be pushed through the grid. We are presently repeating efforts to use this retained bed by agitating it with preheated fluidizing air, immediately after loading. Because of the hygroscopic nature of this retained chemical bed, it may be necessary to use fresh sand for startup bed material.

January 9, 1970

After encountering many minor problems which have been magnified by the inclement weather, tne incinerator was started up on Wednesday. Operations appear to be lining out helping to establish that operation under adverse winter conditions is feasible. We plan to maintain operations as long as adequate sludge feed is available to help slow up excess spent caustic inventory.

January 19, 1970

Operation of the incinerator has continued to be smooth. The bed unloading screw conveyor is down for repairs so it has been necessary to unload the hot bed directly into a skip box. Feed to the incinerator has been kept at less than maximum to reduce the amount of bed to be dumped in this manner.

January 30, 1970

The incinerator was shut down on Thursday after a record uninterrupted run of over three weeks. Some surging had been noted in the reactor, leading to the conclusion that bed agglomeration was occurring. On inspecting the reactor internals today, it was found that there was only a relatively small buildup of bed material on the transition section of the walls and no agglomeration in the bed. The surging was evidently caused by changing fluidization characteristics of the bed, even though particle size distribution appeared good, and this led to incipient defluidization. Additional experimentation will be required to determine the best way to improve fluidization. Minor repairs are being made to the sludge feed pumps and startup is planned for early next week.

APPENDIX C

Mandan Refinery - November 21, 1969

Determination of Fusion Temperature of Chemical Bed (Incinerator)

General

A sample submitted to the laboratory should be at a temperature which would not cause injury or damage. Before commencing week on the sample, check for hazardous temperature. Wear asbestos gloves when handling samples at test temperature.

Procedure

1. Completely fill six coors 30 L crucibles with chemical bed.

2. Place the crucibles in the large muffle furnace.

3. Set the temperature control to 1100°F. and turn on the heat.

4. With long tongs remove one crucible from the furnace when the set temperature has been reached. Let the crucible cool for two minutes, then pour the contents into a small tin can. If the contents pour out, set the temperature 100°F. higher and repeat step 3.

5. Report the set temperature when the chemical bed fails to flow out of the crucible. If chemical bed still flows at set temperature of 1700°F., discontinue the test and report the result as fusion point 1700°F+.

O. Kraav

Okr/zgr

PARTICLE GROWTH RATE

R = rate of increase in particle diameter, in/hr
t = thickness of deposit, in/hr
A = total area of particles, in^2
a = area of a single particle, in^2
M = total mass of particles, lb
m = mass of a single particle, lb
P = density of material, lb/in^2
D = diameter of particles, in
n = number of particles
W = mass rate of material deposited, lb/hr
V = volume of material deposited, in^3/hr

For a single particle, the ratio of surface area to mass is

$$\frac{a}{m} = \frac{\pi D^2}{P \frac{\pi}{6} D^3} \tag{1}$$

For the entire fluid bed of n particles

$$\frac{A}{M} = \frac{6 n \pi D^2}{n P \pi D^3} \tag{2}$$

$$A = \frac{6 M n \pi D^2}{n P \pi D^3} \tag{3}$$

$$A = \frac{6 M}{P D} \tag{4}$$

The Volume rate of material deposited is

$$V = \frac{W}{P} \tag{5}$$

The thickness rate of deposit is

$$t = \frac{V}{A} \tag{6}$$

The diameter of each sphere increases by 2 t

$$R = 2t \tag{7}$$

PARTICLE GROWTH RATE

Substitute the value of t from equation (6)

$$R = 2\frac{V}{A} \qquad (8)$$

Substitute the value of V from equation (5)

$$R = 2\frac{W/P}{A} \qquad (9)$$

$$R = 2\,W/PA$$

Substitute the value of A from equation (4)

$$R = \frac{2\,W}{P\,(6M/PD)}$$

$$R = 2\,W\,P\,D\,/\,6\,PM$$

$$P = \frac{WD}{3M}$$

APPENDIX E

FLUID BED DENSITY, DEPTH, AND MASS

ρb = Fluid bed density Lb/Ft^3

H = Fluid bed depth Ft.

h = Distance between partial bed differential taps ft.

Δp = Partial bed differential, in H_2O

ΔP = Total bed differential, in H_2O

A = Cross sectional area of Bed Ft^2

M = Pounds of Bed

BED DENSITY

$$\frac{\left(\Delta p,\ \text{in } H_2O\right)\ \left(144,\ in^2/ft^2\right)}{\left(27.7,\ \frac{\text{in } H_2O}{Lb/in^2}\right)\left(\ h,\ ft\right)} = \rho b,\ Lb/Ft^3$$

When h = 2.5 ft (30 in)

$$\frac{\Delta p\ \ (144)}{(27.7)\ (2.5)} = \rho b$$

$$\rho b = 2.08\ \Delta p$$

FLUID BED DENSITY, DEPTH, AND MASS

BED DEPTH

$$\frac{\Delta P\, n}{\Delta p} = H$$

When $h = 2.5$ ft (30 in)

$$H = 2.5\ \Delta P / \Delta p$$

Depth to lower differential pressure tap.

BED MASS

$$A\, H\, \rho b = M$$

When $A = 23.76\ ft^2$

$$(23.76)(2.5\ \Delta P / \Delta p)\ (2.08\ \Delta p) = M$$

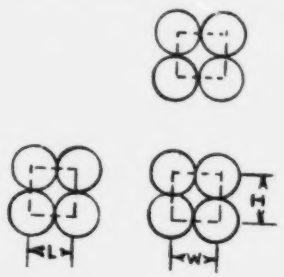
$$M = 123.55\ \Delta P$$

APPENDIX F

APPARENT BULK DENSITY OF UNIFORM SPHERES

ABD = Apparent bulk density
ρ_p = Particle density
V = Volume
L = Length
W = Width
H = Height
D = Diameter of Spheres

Loose Packing



V = Volume occupies by one repetitive configuration

$V = L\ W\ H$

$L = W = H = D$

$V = D^3$

V contains 8, 1/8 spheres or 1 sphere

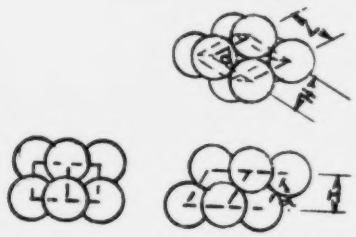
$M = (\rho_p)(\frac{\Pi\ D^3}{6})$

$ABD = M/V$

$$ABD = \frac{\rho_p \frac{\Pi}{6} D^3}{D^3}$$

$ABD = 0.5236\ \rho_p$

Close Packing



V = Volume occupied by one repetitive configuration

$V = L\ H\ H$

$L = D$ $\quad \sin A = H/D$

$A = 60°$ $\quad \sin A = 0.866$

$H/D = 0.866$

$H = 0.866\ D$

$V = (D)(0.866D)(0.866D) = 0.748D^3$

V contains 2, $\frac{0.551}{12.57}$ spheres

6, $\frac{1.911}{12.57}$ spheres

TOTAL 1 sphere

$M = (\rho_p)(\frac{\Pi\ D^3}{6})$

$ABD = M/V$

$$ABD = \frac{\rho_p \frac{\Pi}{6} D^3}{0.748}$$

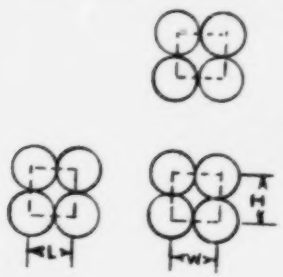
$ABD = 0.700\ \rho_p$

APPENDIX F

APPARENT BULK DENSITY OF UNIFORM SPHERES

ABD	=	Apparent bulk density
ϱ_p	=	Particle density
V	=	Volume
L	=	Length
W	=	Width
H	=	Height
D	=	Diameter of Spheres

Loose Packing



V = Volume occupies by one repetitive configuration

$V = L\ W\ H$

$L = W = H = D$

$V = D^3$

V contains 8, 1/8 spheres or 1 sphere

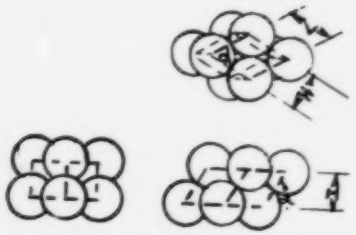
$$M = (\varrho_p)\left(\frac{\Pi\ D^3}{6}\right)$$

$ABD = M/V$

$$ABD = \frac{\varrho_p \frac{\Pi}{6} D^3}{D^3}$$

$ABD = 0.5236\ \varrho_p$

Close Packing



V = Volume occupied by one repetitive configuration

$V = L\ H\ H$

$L = D$ $\quad \sin\Delta = H/D$

$\Delta = 60°\ \sin\Delta = 0.866$

$H/D = 0.866$

$H = 0.866\ D$

$V = (D)(0.866D)(0.866D) = 0.748D^3$

V contains 2, $\frac{0.551}{12.57}$ spheres

6, $\frac{1.911}{12.57}$ spheres

TOTAL 1 sphere

$$M = (\varrho_p)\left(\frac{\Pi\ D^3}{6}\right)$$

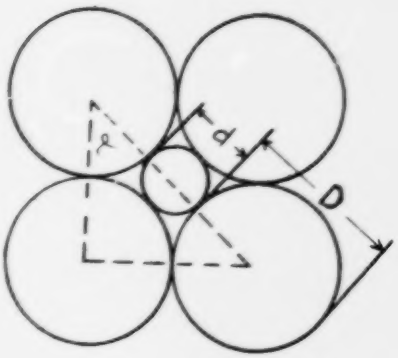
$ABD = M/V$

$$ABD = \frac{\varrho_p \frac{\Pi}{6} D^3}{0.748}$$

$ABD = 0.700\ \varrho_p$

APPARENT BULK DENSITY OF UNIFORM SPHERES

Maximum size small spheres that will fit into spaces among large spheres.



D = Diameter large sphere

d = Diameter small sphere

$$\text{Tan } \alpha = \frac{D}{D} = 1$$

$$\alpha = 45°$$

$$\text{Sin } 45° = 0.7071 = \frac{D}{D + d}$$

$$0.7071D + 0.7071d = D$$

$$0.7071d = D - 7071D$$

$$d = \frac{D(1-0.7071)}{0.7071}$$

$$d = \frac{0.7929\ D}{0.7071}$$

$$d = 0.4142\ D$$

MATERIAL BALANCE

Operation	Test IV	Test VI
Feed Sludge 42 gal/bbl	1.9 bbl/hr @ 8.89 lb/gal (1.98)(42)(8.89) 709.42 lb/hr	2.1 bbl/hr @ 15.52 lb/gal (2.1)(42)(15.2) = 1368.86 lb/hr
Sludge Composition	Wt % lb/hr Oil 31.6 224.17 Water 56.4 400.12 Sediment 12.0 85.13 100.0 709.42	Wt % lb/hr Oil 22.2 303.89 Water 30.8 421.61 Sediment 47.0 643.36 100.0 1368.86
Sediment to Ash	(85.13 lb/hr)(86.2%/100) 73.38 lb/hr	(643.36 lb/hr)(75.8%/100)= 487.67 lb/hr
Water From Combustion	C/H in Oil 6.88 Free Oil in Sludge 224.17 Oil in Sed 85.13-73.38 = 11.75 235.92 lb/hr	C/H in Oil 8.20 Free Oil in Sludge 303.89 Oil in Sed 643.36-487.67 155.69 459.58 lb/hr
C/H = r H + C = Oil C = hr H + hr = Oil H (1 + r) = Oil H = Oil/(1+r) H^2 = 2 11.11% O = 16 88.89% 18 100.00%	C/H = 6.88 6.88 + 1 = 7.88 235.92 / 7.88 = 29.93 lb/hr H^2 29.93 / 0.111 = 269.37 lb/hr H^2O Prod.	C/H = 8.20 8.20 + 1 = 9.20 459.58 / 9.20 = 49.95 lb/hr H^2 49.95 / 0.111 = 449.55 lb/hr H^2O Prod.
Water From Sludge	Free Water 400.12 Combustion Water 269.37 Total 669.49 lb/hr	Free Water 421.61 Combustion Water 449.55 Total 871.16 lb/hr

Operation	Test IV	Test VI
Spent Caustic	1.1 Bbl/Hr @ 10.05 lb/Gal	2.0 Bbl/Hr @ 9.0/ Lb/Gal
42 Gal/Bbl	(1.1)(42)(10.05) = 464.31	(2.0)(42)(9.07) = 761.88
Caustic Composition	Wt % / Lb/Hr	Wt % / Lb/Hr
	Water 62.4 / 289.73	Water 71.6 / 545.52
	NaOH 8.2 / 38.07	NaOH 6.1 / 46.47
D.S.=Dissolved Solids	NO_2CO_3 10.2 / 47.36	NO_2CO_3 7.7 / 58.66
	Other DS 19.2 / 89.15	Other DS 14.6 / 111.23
	100.0 / 464.31	100.0 / 761.88
Water From NaOH Dehydration $2\ NaOH \rightarrow Na_2O + H_2O$ Na=23 H_2= 2 O =16 O =16 H = 1 18 40 X2 X1 80 18	$\frac{(38.07)(18)}{(80)}$=8.56 Lb/Hr Water	$\frac{(46.47)(18)}{(80)}$=10.46 Lb/Hr Water
	Lb/Hr	Lb/Hr
	Free Water 289.73	Free Water 545.52
	Dehydration water 8.56	Dehydration water 10.46
	298.29	555.98
Total Water From Feed to Combustion Gasses	From Sludge 669.49 From Spt Caustic 298.29 967.78	From Sludge 871.16 From Spt Caustic 555.98 1427.14

Operation	Test IV	Test VI
Sulfate Produced	Sludge 709.42 Lb/Hr @ ∧01.25 Wt % S 8.87 Lb/Hr S	Sludge 1368.86 Lb/Hr @ ∧ 00.56 Wt % S 7.66 Lb/Hr S
	Caustic 464.31 Lb/Hr @ ∧ 03.4 Wt % S 15.78 Lb/Hr S	Caustic 761.88 Lb/Hr @ ∧ 02.9 Wt % S 22.09 Lb/Hr S
	Total S = 24.65 Lb/Hr	Total S = 29.75 Lb/Hr
Wt % Na_2 46 32.4 S 32 22.5 O_4 64 45.1 142 100.	24.65/0.225 = 109.55 Lb/Hr Na_2SO_4	29.75/0.225 = 132.22 Lb/Hr Na_2SO_4
Na_2O Required	(109.55)(0.4366) = 47.83 Lb/Hr Na_2O	(132.22)(0.4366) = 57.73 Lb/Hr Na_2O
Na_2 = 46 O = 16 62 / 142 = 0.4366	38.07 Lb/Hr NaOH in Spent Caustic -8.56 Lb/Hr H_2O in NaOH 29.51 Lb/Hr Na_2O from NaOH	46.47 Lb/Hr NaOH in Spent Caustic -10.46 Lb/Hr H_2O in NaOH 36.01 Lb/Hr Na_2O from NaOH
	47.83 -29.51 18.32 Lb/Hr Na_2O from Na_2CO_3	57.73 -36.01 21.72 Lb/Hr Na_2O from Na_2CO_3
$Na_2CO_3 \rightarrow Na_2O + CO_2$	18.32/.5849 = 31.32 Lb/Hr Na_2CO_3 Used	21.72/0.5849 = 37.13 Lb/Hr Na_2CO_3 Used
$\frac{Na_2O}{Na_2CO_3} = \frac{62}{106} = 0.5849$		
$\frac{CO_2}{Na_2CO_3} = 0.4151$	(31.32)(0.4151) = 13.00 Lb/Hr CO_2 Released	(37.13)(0.4151) = 15.45 Lb/Hr CO_2 Released
	47.36 -31.32 16.04 Lb/Hr Na_2CO_3 to Ash	58.66 -37.13 21.53 Lb/Hr Na_2CO_3 to Ash

Operation	Test IV	Test VI
Ash From Feed	Ash From Sludge 73.38 Lb/Hr Other DS From Caustic 89.15 Na_2SO_4 109.55 Na_2CO_3 16.04 Total Ash 288.12	Ash From Sludge 487.67 Lb/Hr Other DS From Caustic 111.23 Na_2SO_4 132.22 Na_2CO_3 21.53 Total Ash 752.65
Air To Incinerator	Combustion Air 1372 SCFM Atomizing Air 218 SCFM 1590 SCFM	Combustion Air 1463 SCFM Atomizing Air 218 SCFM 1681 SCFM
359 Ft^3/Lb MOL 60 Min/Hr 29 Lb/MOL	$\frac{(1590)(60)}{(359)} = 265.74$ MOL/Hr $(265.74)(29) = 7706.46$ Lb/Hr Air	$\frac{(1680)(60)}{(359)} = 280.95$ MOL/Hr $(280.95)(29) = 8147.55$ Lb/Hr Air
Orsat Dry Gas C_2 12 O_2 32 44	MOL/Hr CO_2 = 6.2 15.53X44= 683.32 O_2 =10.0 25.05X32= 801.60 N_2 =83.8 209.93X28=5878.04 100. EDG = 7462.96 (265.74 MOL/Hr Air) (0.79) = 209.93 MOL/Hr N_2 $\frac{209.93}{0.838} = 250.51 \frac{MOL}{Hr}$ Dry Gas	MOL/Hr CO_2 = 7.0 18.94X44= 833.36 O_2 = 11.0 29.77X32= 952.64 N_2 = 82.0 221.95X28= 6214.60 100. 8000.60 (280.95 MOL/Hr Air) (0.79) = 221.95 MOL/Hr N_2 $\frac{221.95}{.82} = 270.67 \frac{MOL}{Hr}$ Dry Gas

Operation	Test IV		Test VI	
		Pounds		Pounds
Input	Sludge	709.42	Sludge	1368.86
	Spent Caustic	464.31	Spent Caustic	761.88
	Air	7706.46	Air	8147.55
	Total	8880.19	Total	10278.29
Output	Dry Gas	7462.96	Dry Gas	8000.60
	H_2O	967.78	H_2O	1427.14
	Ash	288.12	Ash	752.65
	Total	8718.86	Total	10180.39
Difference		161.33		97.90
Increase H_2O For Balance	Dry Gas	7462.96	Dry Gas	8000.60
	H_2O	1129.11	H_2O	1525.04
	Ash	288.12	Ash	752.65
	Total	8880.19	Total	10278.29
	Decrease C/H Estimate From 6.88 to 3.93		Decrease C/H Estimate From 8.20 to 6.56	

EFFECT OF ADDITIONS OF FINE MATERIAL TO THE FLUIDIZED BED

The weight mean particle diameter of the bed is increased by the loss of fine material and reduced by the introduction of fine material with the sludge. When excessive material is lost through the cyclone the effect of the introduction of fine material with the sludge is nullified. However, fine material from some external source can be added to the bed so that there is a net effective addition of ΔM_i with a weight mean particle size D_i.

The weight mean particle size in the bed after an addition of ΔM_i with a size D_i would be

$$\frac{MD + \Delta M_i D_i}{M + \Delta M_i} \qquad (1)$$

The change in weight mean particle size in the bed would be

$$\Delta D = \frac{MD + \Delta M_i D_i}{M + \Delta M_i} - D \qquad (2)$$

With small incremental changes

$$dD = \frac{MD + dM_i D_i}{M + dM_i} - D \qquad (3)$$

But the rate of change in particle size due to particle size growth from the deposition of dissolved solids is

$$R = \frac{WD}{3M} \quad \text{(from Appendix D)} \qquad (4)$$

which can be represented by

$$dD = \frac{dM_S\, D}{3M} \qquad (5)$$

And when the two rate of diameter changes are equal and opposite, an equilibrium bed particle size is reached

$$\left[\frac{dM_S D}{3M}\right] + \left[\frac{MD + dM_i D_i}{M + dM_i} - D\right] = 0 \qquad (6)$$

$$D = \frac{-3D_i M\, dM_i}{M\, dM_S + dM_i\, dM_S - 3M\, dM_i} \qquad (7)$$

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APPENDIX H

Let $F = \frac{dM_i}{dM_s}$ (8)

Then $dM_s = \frac{dM_i}{F}$ (9)

And $D = \frac{-3D_i M_d M_i}{M\frac{dM_i}{F} + \frac{(dM_i)^2}{F} - 3M_d M_i}$ (10)

$DM\frac{dM_i}{F} + D\frac{(dM_i)^2}{F} - 3DMd^{M}i = -3D_i M\, d^{M}i$ (11)

$DM\frac{dM_i}{F} + D\frac{(d^{M}i)^2}{F} = 3DM\, d^{M}i - 3D_i M_d M_i$ (12)

$D\frac{d^{M}i}{F}(M + d^{M}i) = 3M_d M_i(D-D_i)$ (13)

$\frac{D\, d^{M}i(M + d^{M}i)}{3M\, d^{M}i(D - D_i)} = F$ (14)

$\frac{D(M + dM_i)}{3M(D-D_i)} = F$ (15)

Since $d^{M}i$ is very small, it may be omitted.

$\frac{DM}{3M(D-D_i)} = F$ (16)

$F = \frac{D}{3(D-D_i)}$ (17)

Substitute $\frac{d^{M}i}{d^{M}s}$ for F

$\frac{d^{M}i}{d^{M}s} = \frac{D}{3(D-D1)}$ (18)

Equation (18) reveals that there is a real value of $\frac{dMi}{dMs}$ which will produce an equilibrium bed particle size D so long as D_1 is less than D and greater than 0.

The curve in Figure 11 represents the relationship between real values of $\frac{dMi}{dMs}$ and $\overline{D}$. When designing a system, a practical mean bed particle diameter ($\overline{D}$) may be estimated based on dissolved solids in the feed, cyclone losses, insoluble particulate matter in the sludge, and caustic and reasonable fines additions. The design superficial space velocity is then fixed to that velocity necessary to fluidize a bed with the estimated mean particle size ($\overline{D}$).

When operating the unit, space velocity changes with conditions of feed and it is not entirely under the control of the operator. For example, with a system designed to operate with a bed temperature of 1400°F, it might be found that (due to the fusion temperature of the bed material) the temperature must be reduced to 1200°F. This space velocity would be less than design. The inability of the operator to adjust space velocity at will necessitates his being provided with a means by which the mean particle size can be adjusted to produce a properly fluidized bed. This adjustment may be provided by a system for introducing supplemental fine material to the bed or by controlled attrition.

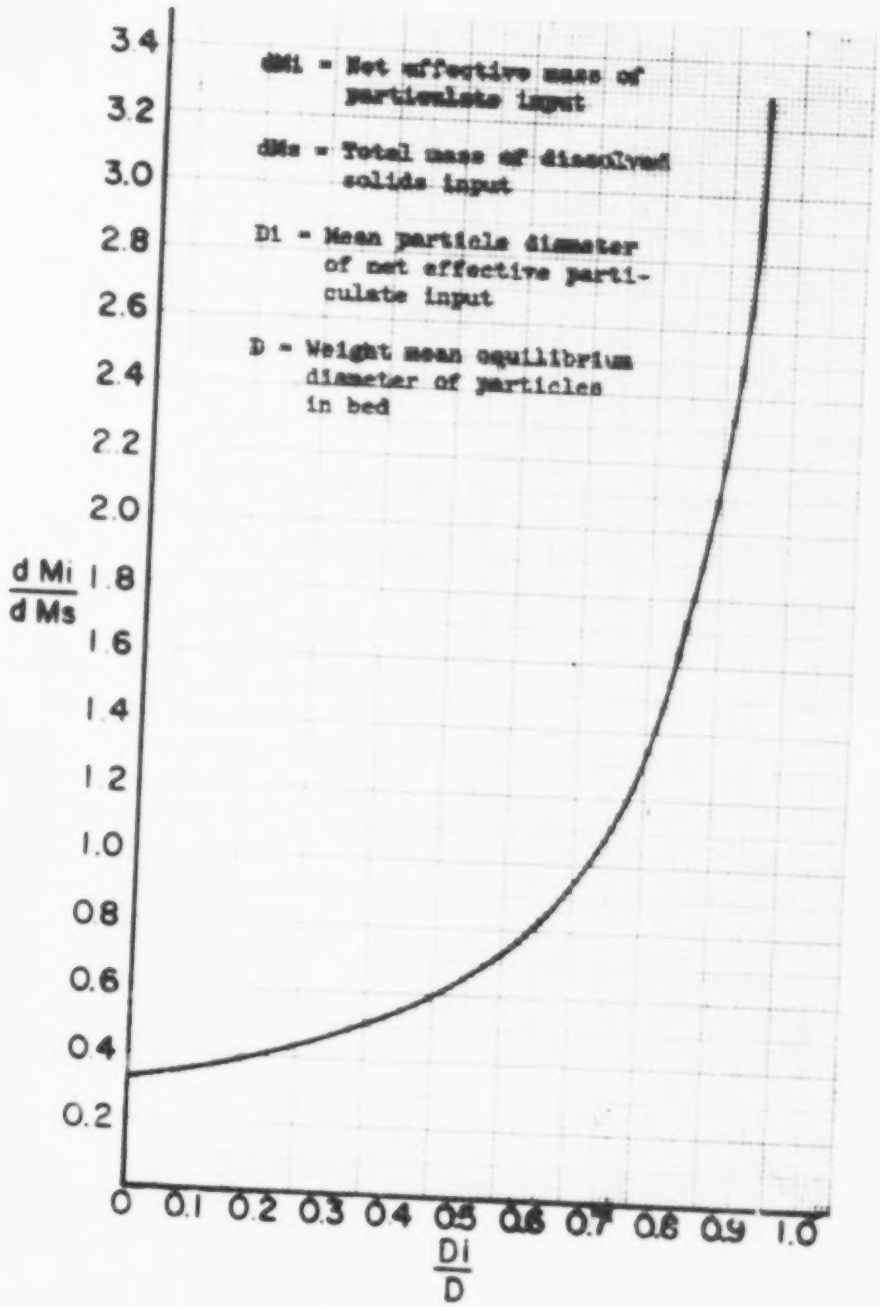

FIG 11 $\frac{Di}{D}$ vs $\frac{d\,Mi}{d\,Ms}$

BIBLIOGRAPHIC:
American Oil Company, "Fluid-Bed Incineration of Petroleum Refinery Wastes", Finai Report FWQA Grant No. WPRD 215-01-68, May 23, 1968 (Project No. 120-50-EKT)

ABSTRACT
The applicability of the Fluid-Bed Incineration process for the disposal of petroleum refinery generated spent caustic and oily sludge in a commercial scale unit has been demonstrated.

Operating problems have been studied. Design and operating procedural changes are suggested.

ACCESSION NO.

KEY WORDS:

Incineration

Fluid-Bed

Spent Caustic

Oily Sludge

Fluidization of Solids

Apparent Bulk Density

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Fluidization of Solids

Apparent Bulk Density

1 Accession Number	2 Subject Field & Group	SELECTED WATER RESOURCES ABSTRACTS Input Transaction Form

5 Organization: American Oil Company

6 Title: Fluid Bed Incineration of Petroleum Refinery Wastes

10 Author(s)	11 Date	12 Pages	5 Contract Number
Herbert E. Simons	11-20-70	107	WPRD 215-01-68
	16 Project Number: 12050 EKT		21 Note

22 Citation

23 Descriptors(Starred First)

25 Identifiers(Starred First)

27 Abstract

The applicability of the fluid bed incineration process for the disposal of petroleum refinery generated spent caustic and oily sludge in a commercial scale unit has been demonstrated.

Operating problems have been studied. Design and operating procedural changes are suggested.

The major process limitation stems from the loss of bed fluidity due to high particle size growth rate. Particle size growth rate is directly proportional to the particle diameter and rate of dissolved solid material charged and inversely proportional to the mass of material in the bed. The average particle diameter can be controlled by (1) collecting and continuously returning fine material to the bed (2) utilizing an effective attriting system and (3) limiting superficial space velocity to avoid elutriation of fines. This report was submitted in fulfillment of Project #12050 EKT under the partial sponsorship of the Federal Water Quality Administration.

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WR:162 (Rev. Oct. 1966)
WRSIC

Send To: Water Resources Scientific Information Center, U. S. Department of the Interior Washington, D.C. 20242

★U.S. GOVERNMENT PRINTING OFFICE: 1971-423-754/1164

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